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EVALUATION OF THE FLUIDIZED-BED COMBUSTION PROCESS
VOLUME IV. FLUIDIZED-BED OIL GASIFICATION/ DESULFURIZATION

WESTINGHOUSE RESEARCH LABORATORIES

PREPARED FOR
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EVALUATION OF THE FLUIDIZED-BED COMBUSTION PROCESS

VOLUME IV - FLUIDIZED-BED OIL GASIFICATION/DESULFURIZATION

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December 1973

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PREFACE

The Office of Research and Development (ORD) of the United States Environmental Protection Agency (EPA) has organized and is sponsoring a fluidized bed fuel processing program. Its purpose is to develop and demonstrate new methods for utilizing fossil fuels -- particularly coal and oil -- in utility power plants. These methods should:

- Meet environmental goals for SO_2, NO_x, ash, smoke emissions, and wastes
- Compete economically with alternative means for meeting these abatement goals.

Westinghouse Research, under contract to the Office of Research and Development (ORD) of the Environmental Protection Agency (EPA), is carrying out a study to evaluate and develop fluidized bed combustion and oil gasification in air pollution abatement. The goals of this work are:

- To identify which fluidized processes might be economically employed in utility power plants or industrial boilers to reduce SO_2, particulate, and NO_x emissions
- To assist in planning and implementing a program to develop the fluidized bed systems deemed effective in air pollution control and economical in steam/power production.

Tasks in the evaluation of fluidized bed combustion set forth by EPA which have been completed at Westinghouse under a previous contract are:

- To search the technical and patent literature in fluidized bed combustion, to canvass commercial organizations with expertise pertaining to this field, and to survey the market for industrial boilers and utility power systems

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- To design a fluidized bed industrial boiler and two fluidized bed utility power systems -- one 300 MW capacity, the other 600 MW; and to provide performance and cost projections for the equipment designed
- To provide technical consultation and assistance on the ORD fluidized bed fuel processing program, including both combustion and gasification
- To conceptualize a development plant which will prove the configurational or operating features of the fluidized bed boiler designs not confirmed by prior art or experience
- To assess the effectiveness and economics of an atmospheric-pressure, fluidized bed oil gasification-combustion system and to aid in planning for a demonstration installation of such a system.

The results of these surveys, designs, evaluations, and comparisons were published in a three-volume report, "Evaluation of the Fluidized Bed Combustion Process," in November 1971 under contract No. CPA 70-9.

These results provided the basis for the work performed from July 1971 to May 1973 which is described in this four-volume report. Tasks in the evaluation of the fluidized bed combustion process set forth by EPA under contract 68-02-0217 have focused on the development of pressurized fluidized bed combustion for power generation and fluidized bed oil gasification for power generation. These tasks have included:

- Extensive process evaluation studies of the pressurized fluidized bed combustion system for power generation. These investigations predict the sensitivity of operating and design parameters selected for the base power plant design on plant economics; provide additional experimental data on sulfur removal and sorbent regeneration using limestones and dolomites; project economics and performance for various sulfur removal systems, both regenerative/sulfur recovery and once-through sorbent; establish a plant operation and control philosophy and evaluate alternative pressurized fluid bed combustion concepts, both economic and performance.

- Preparation of preliminary plans and a cost estimate for a 30 MW (equivalent) pressurized fluidized bed combustion boiler development plant. The design provides sufficient detail to locate a suitable plant site and to obtain a fixed cost bid for the preparation of detailed plans. The plant will provide the capability to study effectively the remaining technical problems to achieve the greatest potential for reducing emissions and to achieve economical power generation.

- Identification of a project team to demonstrate fluidized bed oil gasification/desulfurization for power generation. A cooperating utility -- New England Electric System (NEES) -- has been identified to carry out a 50 MW demonstration plant program. Further process evaluation has been carried out and experimental data obtained on sulfur removal and spent stone disposal.

- Evaluation of pressurized oil gasification for combined cycle power generation. Oil gasification process concepts and options have been reviewed, material and energy balances projected, performance projected, and capital and energy costs estimated.

- Provision of technical consultation and assistance on the ORD fluidized bed fuel processing program, including both combustion and gasification processes. Technical and economic comparisons have been carried out on various fluidized bed fuel processing systems and various conventional means of steam/power generation.

This volume (Volume IV) contains further evaluation of the atmospheric pressure fluidized bed oil gasification system, the development of a 50 MW demonstration plant team and program, and a technical and economic evaluation of pressurized oil gasification for combined cycle power generation. Volume I contains new data on pressurized fluid bed combustion and an evaluation of the current technology state. Volume II contains the appendices to Volume I. Volume III presents the preliminary design of a 30 MW (equivalent) pressurized fluid bed boiler development plant.

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CONVERSION FACTORS -- ENGLISH TO METRIC UNITS

	English System	Metric Equivalent
Length	in	2.54 cm
	ft	0.305 m
Area	in^2	6.45 cm^2
	ft^2	0.0930 m^2
Volume	in^3	16.39 cm^3
	ft^3	28.32 l
	gal	3.785 l
Mass	oz	28.35 gm
	lb	453.6 gm
	ton	907.2 kg
Pressure	lb/in^2	51.70 mm Hg
	in H_2O	1.865 mm Hg
Temperature	°F	1.8 (°C) + 32
	°R	1.8 °K
Energy	Btu	252 cal
Power	Btu/min	252 cal/min

1. SUMMARY

1.1 INTRODUCTION

The production of low-Btu gas from oil is attractive to the electric utility industry because the supply of natural gas and clean fuel oil is being depleted, their cost is prohibitive, and the technologies for removing sulfur dioxide from stack gases are proving to be expensive. Fluidized bed oil gasification will offer an effective and economical method of utilizing relatively abundant, low-cost, high-sulfur oil or low-grade liquid fuels for producing a low-sulfur fuel gas suitable for power generation in a conventional steam power plant or combined cycle power plant. Atmospheric-pressure fluidized bed oil gasification (which was originally developed by Esso Petroleum Co.[England], subsequently funded by the Environmental Protection Agency [EPA], and then evaluated by Westinghouse under contract to EPA) can reduce sulfur emissions by 90 to 95% and minimize NO_x formation from oil-fired utility power plants.

Fluidized bed oil gasification can be applied to power generation to produce a clean, low-Btu fuel gas -- 200 to 500 Btu/scf. In a fluidized bed gasifier, oil is added to a simple, open, fluidized bed of limestone or dolomite with sufficient air -- 15 to 25% of stoichiometric -- to maintain the bed at ∿1600°F and react with the oil to produce a fuel gas. The limestone or dolomite removes sulfur from the fuel gas during the gasification process.

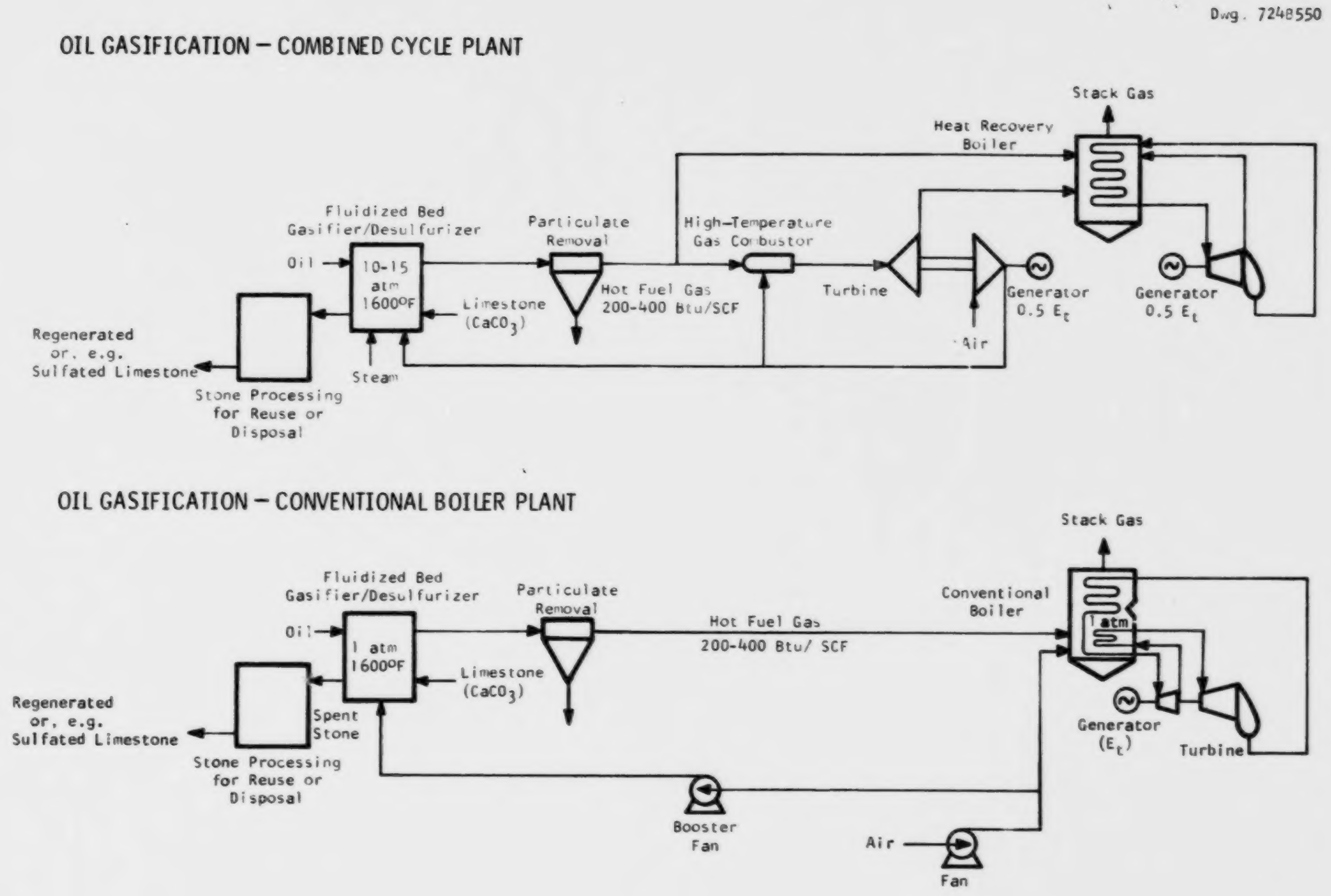
The process can be operated as a once-through limestone sorbent system, a sorbent regeneration/sulfur recovery system, or a sorbent regeneration system without sulfur recovery, by capturing the sulfur-rich gas from the regenerator with the spent stone. The spent stone from each system alternative can be processed to minimize the environmental impact of the waste stone for disposal or to provide a material for potential market utilization.

Gasification of oil can be carried out at either atmospheric or elevated pressure. Operating at atmospheric pressure, a fluidized bed gasifier provides clean fuel to a boiler. Operating at elevated pressure, a gasifier provides clean fuel to a combined cycle gas- and steam-turbine power plant. The potential for high efficiency and low capital cost of such a plant makes this system attractive. Power cycle schematics for these two alternatives are shown in Figure 1.

Fluidized bed oil gasification provides the technology for utilizing high-sulfur residua and asphalts for clean power generation. These fuels cannot be utilized by conventional methods to provide clean, reliable power at low cost. Although many fuel oil desulfurization facilities are being built, it does not appear that adequate quantities of low-sulfur oil will be available to meet future demand. The development of processes to utilize these high-sulfur, low-grade fuels is necessary to relieve the energy shortage caused by both a lag in nuclear energy and coal utilization technologies and increasing pollution control requirements. The simplicity of the fluidized bed oil gasification process provides a new technology which is near commercial development and can relieve this energy shortage.

1.2 TECHNOLOGY ASSESSMENT

The atmospheric-pressure operation has been demonstrated by Esso Petroleum Co. Ltd. on a 1 MW pilot-plant scale. The pilot-plant experience demonstrates the ability to control SO_2 and NO_x. Sulfur removal up to 95% can be achieved. Nitrogen oxide emissions of 150 ppm appear possible. Particulate emissions from the fluid bed unit can be removed to achieve proposed standards. The pressurized system will reduce thermal emissions by 40 to 50%. In addition, plant availability should increase because of the ability of the fluidized bed gasification process to remove vanadium, reduce sodium, and reduce the SO_3 levels in the boiler, thereby,in turn,reducing corrosion and fouling.



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OIL GASIFICATION – COMBINED CYCLE PLANT

OIL GASIFICATION – CONVENTIONAL BOILER PLANT

Figure 1—Fluidized Bed Oil Gasification for Power Generation from Fossil Fuels

Westinghouse, under contract to the Office of Research and Development, EPA, carried out a design study and evaluation of fluidized bed oil gasification/desulfurization at atmospheric pressure for power generation.[1] The atmospheric pressure process can be used as a retrofit on an existing boiler or as a design feature on a new boiler. Both new and retrofit concepts provide advantages over stack gas scrubbing pollution control. Capital cost of a retrofit, once-through oil gasification system may be 50 to 70% less than a retrofit stack gas cleaning system. The capital cost of new gasification systems may be 20 to 50% less than that of a stack gas cleaning system. Total plant cost savings may be 4%. Fuel adder costs will be between 20 to 50% less than stack gas cleaning, low-sulfur oil, or desulfurized oil. Total energy cost savings may be as high as 3%.

The utilization of the pressurized fluid bed oil gasification process with current combined cycle power generation technology provides higher plant efficiencies than can be obtained with conventional power plants and controls pollution to low levels. Pressurized fluidized bed oil gasification offers the potential for achieving power plant efficiencies greater than 50% by going to higher gas-turbine inlet temperatures. A second generation plant may achieve 50% efficiency. The fluidized bed gasification process enables the normally unusable fuels such as asphalt to be consumed in combined cycle plants which presently require clean distillate fuels or natural gas, permitting these valuable fuels to be conserved for other purposes. Energy cost reductions of 15 to 20% compared to conventional plants using wet scrubbing for pollution control or combined cycle plants using distillate fuels are projected.

Both atmospheric pressure and pressurized operations may use once-through operation or regeneration of the limestone or dolomite sorbent. The waste material problem is minimized, since the sorbent waste can be maintained dry, potentially suitable for other uses.

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[1] "Evaluation of the Fluidized Bed Combustion Process." Final Report, OAP Contract 70-9, Westinghouse Research Laboratories, Nov.1971, (three volumes, NTIS PB 211-494, PB 212-916, PB 213-152).

Fluidized bed oil gasification can be utilized for industrial fuel gas and/or steam generation. Industrial applications of fluidized bed oil gasification offer economical and environmental advantages over stack gas cleaning if clean fuels are not available. The once-through operating mode appears most attractive for industrial applications.

1.3 ENVIRONMENTAL IMPACT

Environmental factors for power generation by oil gasification with a conventional plant, or with a combined steam- and gas-turbine plant, are compared with the environmental effects of a conventional oil-fired power plant using limestone wet scrubbing in Table 1. Included in the table are the plant capital investment and energy cost, which may range over some set of values depending on the specific plant conditions (new plant or retrofit, plant location, etc.). For instance, the capital cost of a new plant with limestone wet scrubbing may range from \$25 to 40/kw, while a retrofit plant may range from \$40 to 70/kw. Each of the processes has some advantages. The major advantage of the limestone wet scrubbing process is the status of its development. The oil gasification processes have cost advantages and potentially lower waste and emission levels.

The spent stone from the fluidized bed oil gasification/desulfurization process is dry and granular (-1/4 inch). The composition of the spent stone is primarily calcium oxide and calcium sulfide with small amounts of calcium sulfate -- the spent stone from regenerative operation containing ∿ 1 to 2% CaS, and from once-through operation a projection of 30 to 80% CaS. The disposal of this solid may be accomplished by a variety of methods; the preferred method will reflect local circumstances. Several processing alternatives are currently under study to prepare the stone for landfill and/or market. Both dry and wet processes are being developed to convert the CaO and CaS to an environmentally acceptable form for disposal. Dry sulfation followed by flue gas carbonation and sintering at high temperature are examples of dry processing systems; slurry carbonation is an example of a wet process being developed. Development of alternative stone disposal processes continues.

TABLE 1

ENVIRONMENTAL IMPACT COMPARISON

	Conventional Oil-Fired Power Plant with Limestone Wet Scrubbing	Conventional Oil-Fired Plant with Atm-Pressure Oil Gasification		Combined Cycle Power Plant with Pressurized Oil Gasification	
		Regenerative	Once-through	Regenerative	Once-through
Plant Capital Investment (\$/kw)	325 (scrubber cost of \$50/kw)	310	300	180	160
Plant Energy Cost (mills/kWh)	13.1	12.7	12.5	9.7	9.2
Limestone Waste (tons/day)	480 (sludge)	90-180 (dry)	500-800 (dry,$CaSO_4$)	90-180 (dry)	500-775 (dry)
Water Usage (tons/day)	500 (sludge)	--	--	--	--
SO_2 Emission (lb/MM Btu)	0.35	0.35	0.35	0.35	0.35
NO_x Emission (lb/MM Btu)	0.8	0.16	0.16	0.16	0.16
Particulate Emission (lb/MM Btu)	0.02	0.02	0.02	0.02	0.02
Thermal Emission (% of conventional plant)	--	∿ same	∿ same	∿ 50	∿ 50
Auxiliary Power & Fuel (% of plant input)	4.0	2.5	2.5	7.5	5.0
Plant Heat Rate (Btu/kWh)	9600	9500	9500	9500	9200
Land Requirement (compared to a conventional plant)	large area for disposal pond	∿ same	∿ same	more compact power plant	

Basis: 635 MW power plant
1973 costs
Capitalization at 15%
Residual oil at 45¢/MM Btu
Limestone purchase plus disposal at \$6/ton
3 wt % sulfur in fuel oil
90% sulfur removal
Gasification at air/fuel ratio of 20% of stoichiometric
Limestone make-up rate of 0.5 to 1.0 times stoichiometric for regeneration and 1.5 to 3.0 times stoichiometric for once-through; for wet scrubbing a make-up rate of 1.2 times stoichiometric.
NO_x emission for gasifier based on pilot plant measurements and experience with fluidized bed combustion.
Combined cycle plant based on conventional gas-turbine technology. Advanced gas-turbine conditions can reduce plant costs and total emissions.
Gasification process consists of two independent gasifier modules. No reliability factor is built into the modular design, with each module being capable of 50% of the plant load. Duel fuel capability is included in the plant design.

1.4 DEVELOPMENT

Fluidized bed oil gasification systems -- both atmospheric-pressure and pressurized -- for power generation can be commercially demonstrated by 1977. The technical feasibility of the process, including pollution control, has been demonstrated on a 1 MW pilot plant. Westinghouse has located a utility partner to carry out an atmospheric-pressure demonstration plant program. All utilities visited encouraged support of the development program. New England Power Service Co. of New England Electric System (NEES) has agreed to participate in the atmospheric-pressure demonstration plant program. The program will require approximately three years.

A three-phase demonstration plant program has been conceived:

- Phase I. Preliminary design and cost estimate of demonstration plant installed on an existing boiler plus the subsequent preparation of a bid package and a definitive estimate for Phase II work
- Phase II. Detailed design and construction of demonstration oil gasification process
- Phase III. Developmental operation of the gasification process and integrated power plant.

Westinghouse, as prime contractor to EPA, will subcontract for the engineering services for Phase I. Phase I has two sequential parts:

- A preliminary design and budget estimate of investment for evaluation of the preliminary design, followed by decision to stop or proceed
- Preparation of a bid package and a definitive installed cost estimate.

The bid package is to be adequate for obtaining comparable lump-sum bids for the Phase II work. NEES, as the cooperating utility, in Providence, Rhode Island, will supply technical information on the Manchester Street power plant; provide price and availability information on oil fuels which might be utilized in the plant currently and/or in the future; supply information concerning the load requirements placed on the plant; cooperate in the selection of an engineering firm to prepare the preliminary design and bid package, assist in selecting various options in the design of the system and evaluate, in cooperation with Westinghouse, Esso (England), and EPA, the effectiveness and economy of the oil gasifier/desulfurizer in power generation and pollution abatement.

The design effort will be based on experimental data from Esso (England) on their 1 MW continuous pilot plant, the conceptual design and assessment by Westinghouse, and operating and environmental constraints established by NEES and local regulations. EPA will provide general guidance and funding for Phase I.

If the preliminary design confirms the effectiveness and economics of the system, a proposal will be prepared for Phases II and III. In Phase II the utility would work closely with an engineering firm in the design and installation of the gasifier/desulfurizer system, and share in the costs of the design and installation.

In Phase III the plant would be operated to collect basic data on the operation of the gasifier/desulfurizer system; data would be interpreted and analyzed; and an evaluation of the effectiveness -- technical and economic -- of the process in power generation and pollution abatement would be made.

The development of a pressurized fluidized bed system is not currently supported. If development were initiated immediately, the process could be commercially demonstrated by 1978 at a cost of approximately $\$10 \times 10^6$.

If all the new oil-fired power plants projected for 1980 to 1990 were pressurized fluidized bed oil gasification combined cycle plants operating at 45% efficiency, the energy resource savings would be approximately 50 million bbls of oil per year. This is sufficient fuel for ∿ 6000 MW of generating capacity. The $10 million development cost for the pressurized system is equivalent to the projected capital cost savings on one 200 MW plant. There are ∿ 150 of equivalent sized plants projected for installation between 1980 and 1990.

1.5 CONCLUSIONS

- Atmospheric-pressure fluid bed oil gasification will meet environmental emission standards and will reduce incremental energy costs by up to 50% compared with stack gas cleaning based on system economics and pilot plant results.
- A 50 MW atmospheric-pressure fluid bed oil gasification system demonstration plant program is under way with New England Electric System.
- Pressurized fluid bed oil gasification with combined cycle power generation has the potential to burn high-sulfur residual fuel oils, meet environmental emission standards, and deliver energy at costs 20% less than power generation with clean fuels.

2. ATMOSPHERIC-PRESSURE FLUID BED OIL GASIFICATION

The gasification/desulfurization of residual oil at atmospheric pressure is being evaluated by Westinghouse under contract to the Office of Research and Development (ORD) of the Environmental Protection Agency. The concept is being studied for the purpose of producing, on-site, a low-sulfur fuel gas suitable for power plant utilization in a conventional boiler. Esso Petroleum Company has provided experimental information on the oil gasification/desulfurization process based on small-scale batch fluidized units and a 750 kw continuous unit. Westinghouse has carried out preliminary conceptual design studies to evaluate the commercial process and has located a utility partner with which to carry out a demonstration plant operation.

2.1 PROCESS EVALUATION

2.1.1 Gasification/Desulfurization Concepts

The gasification/desulfurization operation can be accomplished by either the regenerative or the once-through mode. Figure 2 illustrates the major process streams and identifies the basic elements of the two operational modes.

The major components of the regenerative operation are the gasifier and regenerator vessels. The gasifier is an air-fluidized bed of lime operated at 1600°F with substoichiometric air (∿ 20% of stoichiometric). The heavy residual oil is injected into the gasifier vessel, where it cracks and is partially combusted to form a hot, low-sulfur fuel gas. Hydrogen sulfide produced during gasification reacts with the lime to produce a sulfided lime.

$$H_2S + CaO \rightarrow CaS + H_2O \quad .$$

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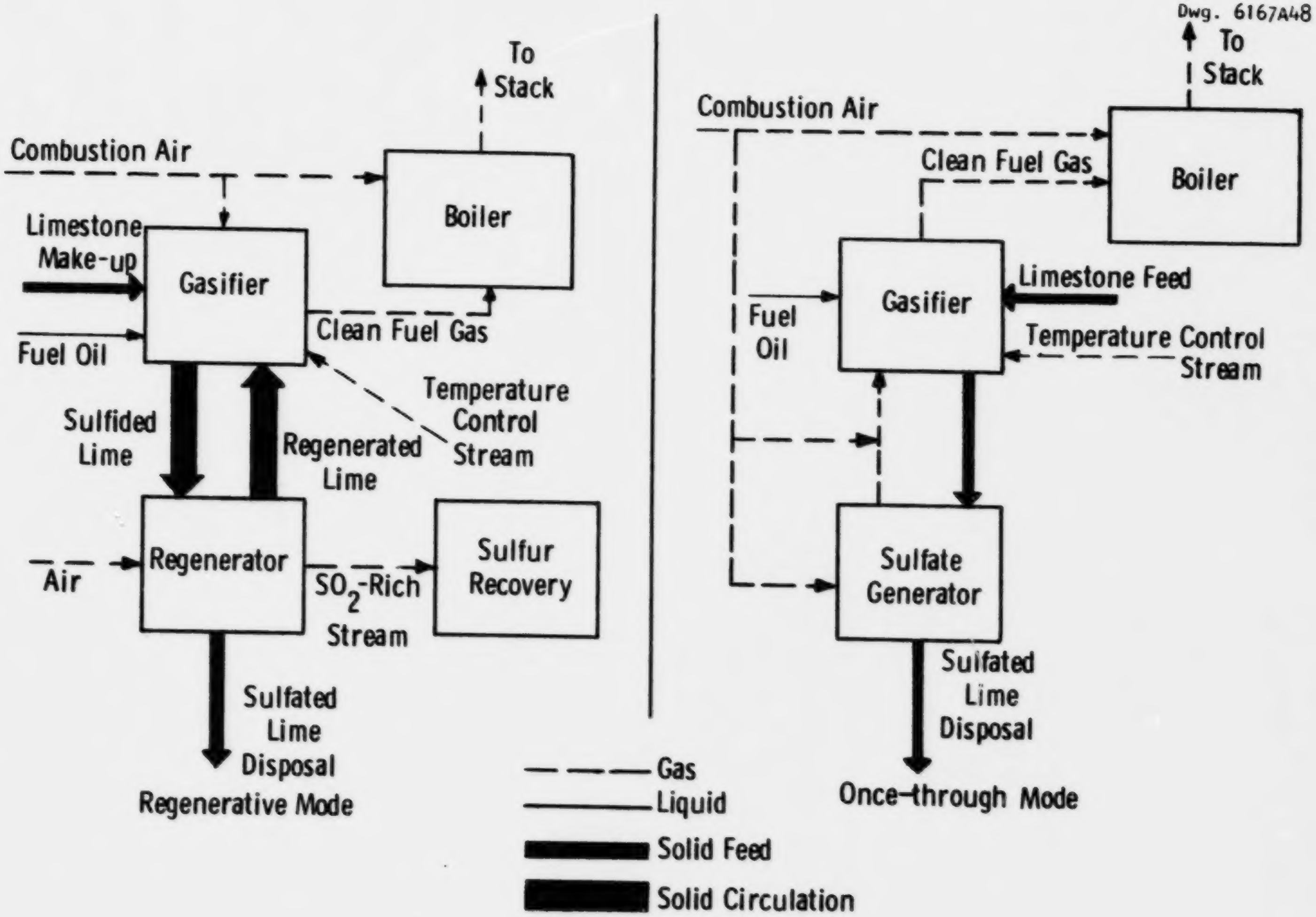

Figure 2—Modes of Operation

The hot fuel gas is transported to the boiler burners where combustion is completed, and the sulfided lime is sent to the regenerator. The regenerator is an air-fluidized vessel operated with a slight excess of air at about 1900°F. Regeneration takes place by reaction of oxygen with the utilized lime to give an SO_2-rich stream (of about 10 mole % SO_2) and a regenerated lime having a slightly decreased activity compared to that of fresh lime.

$$CaS + \frac{3}{2} O_2 \rightarrow CaO + SO_2 \quad .$$

The SO_2 stream is transported to a sulfur recovery system, and the regenerated lime is returned to the gasifier along with an approximately stoichiometric amount of fresh make-up limestone.

The elements of the once-through operation shown in Figure 2 consist of a gasifier vessel and a sulfate generator, or predisposal vessel. The operation of the gasifier for once-through limestone operation is the same as for the regenerative operation. The sulfate generator operates similarly to the regenerator but at a lower temperature ($\sim$ 1500°F) so that the sulfided lime from the gasifier is converted to calcium sulfate rather than calcium oxide. The dry calcium sulfate may be disposed of while the gas stream from the sulfate generator is sent to the gasifier. A limestone additon rate of up to three times that used in the regenerative operation may be necessary in the once-through operation to achieve similar sulfur removal of 90 to 95%. Lower limestone rates may be suitable for once-through operation.

2.1.2 Experimental Work

Under contract to ORD, Esso (England) is carrying out laboratory tests on atmospheric-pressure batch fluidized bed equipment to investigate lime sulfur absorption and lime regeneration operating variables.[1,2] The results of the Esso Petroleum Company experimental program have identified the critical phenomena associated with atmospheric-pressure gasification/desulfurization and have established the probable operating conditions and behavior of a commercial system. The process evaluation has been based mainly on the results of small-scale batch work, which has been carried out in conjunction with the construction and operation of a 750 kw continuous pilot unit. The unit has been operated for a total of more than 600 hours. The information obtained from the batch tests and the continuous unit provide the basis for the design of a gasification/desulfurization retrofit system to be operated as a demonstration unit on a utility boiler. A summary of the experimental results appears in Table 2.

2.1.3 Design

Energy and material balances provide the basic information with which the feasibility of applying the gasification/desulfurization concept as an add-on to an existing boiler, or as a new plant design feature, have been examined. Energy and material balances have been reported.[1] The feasibility of the retrofit concept has been examined in terms of the availability of space in an existing power plant, the modifications necessary to retrofit an existing boiler, and the performance of a modified boiler. The performance of the fluidized bed oil gasifier as a function of the air/fuel ratio is discussed in Appendix A. The gasifier heat generation rate, the gasification system thermal efficiency, and the fuel gas characteristics are considered. The air/fuel ratio and method of gasifier temperature control are shown to be critical factors in determining the process performance. The specifications assumed for the conceptual design of a commercial plant are presented in Table 3.

2.1.3.1 Retrofit Concepts

The feasibility of converting an existing boiler to one which utilizes the fuel from a gasification/desulfurization system depends on a number of factors, many of which will differ from one boiler to the next. The space available for a gasification/desulfurization system in an existing power plant, the modifications necessary to retrofit an existing boiler, and the performance of a retrofit boiler will depend on the specific gasification/desulfurization system design and choice of

TABLE 2

SUMMARY OF EXPERIMENTAL RESULTS

Batch Tests

- Nearly 100% sulfur removal at 1600°F, 20% air/fuel ratio, with limestone make-up rate of 1 mole CaO/mole of sulfur fed
- Nearly 100% of fuel vanadium removed
- About 20% of fuel sodium removed
- Clean product fuel is easily combustible with a luminous, smoke-free flame.
- Regeneration at 1900°F gives a 10% SO_2-gas product and a regenerated lime having reduced activity.

Continuous Unit

- Results indicate superior behavior during continuous operation due to improved temperature control.
    - Lower air/fuel ratios, down to 17% of stoichiometric, have been demonstrated.
    - 90% sulfur removal has been demonstrated.

TABLE 3

ATMOSPHERIC-PRESSURE OIL GASIFICATION SPECIFICATIONS

Process Specifications	Design Variables
Sulfur Removal: 90-95%	Boiler Size: 600 MW
Fuel Oil: 3 wt % sulfur; LHV = 17,700 Btu/lb	Load Factor: 40%, 80%
Limestone: 98.6% CaO yield	Turndown: 4/1

	Operating Variables	
	Regenerative System	Once-through System
Gasifier Temperature	1600°F	1600°F
Regenerator (sulfate generator) Temperature	1900°F	1500°F
Stone Make-up Rate	1 mole CaO/mole sulfur	3 moles CaO/mole sulfur
Air/Fuel Ratio	20% of stoichiometric	20% of stoichiometric
Limestone Utilization	∿ 5 wt % sulfur in bed	∿ 19 wt % sulfur in bed
Fluidization Velocity	8 ft/sec	8 ft/sec
Minimum Fluidization Velocity	3 ft/sec	1 ft/sec
Particle Sizes (avg.)	∿ 2000 μ	∿ 1000 μ
Gasifier Bed Depth (static)	2.5-3.5 ft	3.5-4.0 ft

operating conditions, the location of the gasification/desulfurization system in the plant, the type of boiler (coal-, oil-, or gas-fired), size of boiler, the turndown needed, the boiler load factor, and the specific design features of the boiler.

The gasification system may be placed internal to the boiler (directly beneath) or external to the boiler. Preliminary considerations seem to favor the external retrofit design over the internal retrofit design and the feasibility of retrofitting a coal-fired boiler over the feasibility of retrofitting an oil- or gas-fired boiler. The internal design concept appears to be limited by the available space beneath the boiler and the cost of modifying the boiler. The external design concept should reduce modification costs and should allow greater uniformity and flexibility in system design. Coal-fired boilers have potentially more space near the boiler than oil- or gas-fired boilers and may provide some of the needed solids handling and particulate clean-up equipment. Boiler derate should be of little concern with coal-fired boilers because they are designed to operate with slag on the boiler heat transfer surface. Derate may be a concern with oil- and gas-fired boilers due to a reduction in the flame temperature and redistribution of the heat release within the boiler which may occur with the fuel. Hot fuel gas piping sizes, gasification system space requirements, the effects of using multiple gasifier modules, and the problem of gasifier start-up are considered in Appendix B.

In contrast to boiler retrofit considerations, the feasibility of incorporating a gasification/desulfurization system into a new boiler design will be limited only by the overall economics of the system and the market potential for new oil-fired boilers. The total space occupied by the gasification/desulfurization system will be a small percentage of the total plant volume. Also, because of the flexibility in boiler design, boiler performance will not be affected by the presence of the gasification/desulfurization system.

2.1.3.2 Plant Layout

Figure 3 shows a plant layout for a 600 MW once-through gasification/desulfurization system. The plant consists of two gasification modules, each utilizing air from one of the two existing power plant draft fans.

2.1.4 Evaluation

The design study, coupled with the experimental work of Esso Petroleum Company, points out the technical and economic feasibility of oil gasification/desulfurization as a retrofit SO_2-control system for utility boilers, or as an SO_2-control system to be incorporated into a new boiler design. A market for retrofit and new oil-fired boilers exists.[3]

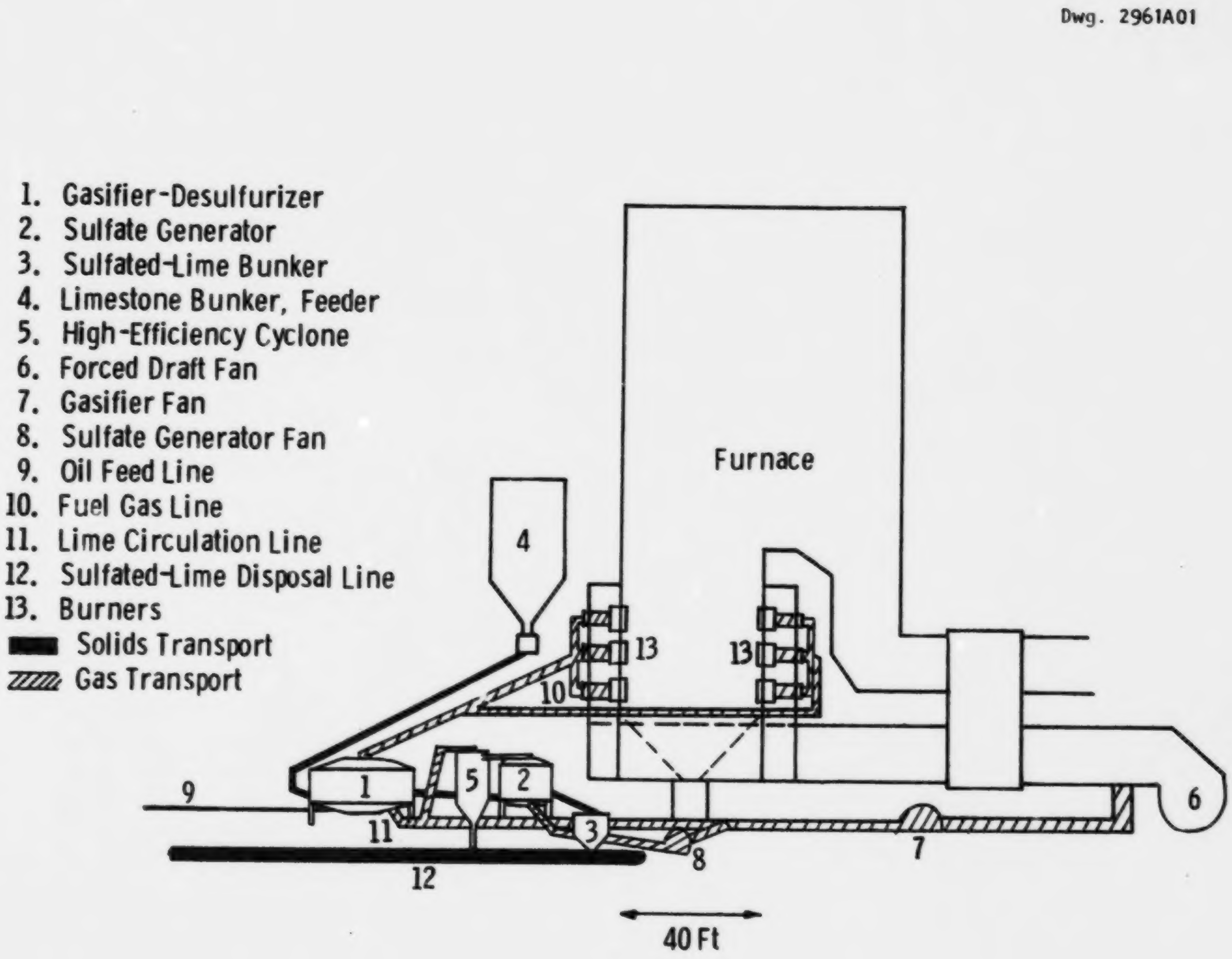

Figure 3—600MW Boiler Retrofit Once-through Design

2.1.4.1 Utility Application

Preliminary investigations indicate that, overall, the once-through operation may be somewhat more attractive to a utility customer than the regenerative operation. Capital investment is reduced with the once-through operation, and, although the limestone feed rate is expected to be up to three times the rate with regenerative operation, the operating costs for once-through operation may be less than those for the regenerative operation. The capital investment and fuel adder costs of regenerative and once-through operations are shown in Appendix C as a function of the sulfur removal efficiency, the limestone cost, the fuel-oil cost, and the plant capacity factor. A complete cost breakdown for once-through and regenerative operation has been presented.[4] Once-through operation has fewer technical problems at this time and is overall a simpler process than regenerative operation.

A comparison is made in Table 4 between atmospheric-pressure oil gasification/desulfurization and the alternative schemes of low-sulfur oil and stack gas cleaning. Capital costs and operating costs are compared for new and retrofit systems. Oil gasification/desulfurization compares favorably with low-sulfur oil and stack gas cleaning, based on preliminary cost estimates. A reduction of about 40% in the capital costs involved in stack gas cleaning is estimated for new and retrofit gasification/desulfurization systems. Operating costs appear to be about the same for once-through stack gas cleaning and regenerative gasification/desulfurization with sulfur recovery. Once-through gasification/desulfurization may reduce the operating costs 30 to 50% as compared to stack gas cleaning. The cost estimate indicates that the operating cost with low-sulfur fuel oil will be about 30 to 50% greater than that for gasification/desulfurization. These conclusions are based on the desulfurization of high-sulfur residual oil (3 wt % sulfur) and may be altered when a lower sulfur oil is considered (1 to 1.5 wt % sulfur). Environmental factors are also compared in Table 4. The low-sulfur oil is advantageous since capital costs are limited to possible boiler modifications necessary when changing from gas or coal to low-sulfur oil. On the other hand, operating costs are higher than those for stack gas cleaning or oil gasification/desulfurization. Capital

TABLE 4

ASSESSMENT OF FLUIDIZED BED OIL GASIFICATION/DESULFURIZATION[a]

	Low-Sulfur Oil	Stack Gas Cleaning	Oil Gasification	
			Regenerative Operation	Once-through Operation
Cost				
Capital, \$/kw				
New	---	25-40	12-15	8-10
Retrofit	b	40-75	22-27	18-22
Fuel Adder, ¢/10^6 Btu				
New	25-35	11-14	9.5-16.0	9-10.5
Retrofit	25-35[b]	14-20	11-18	10.5-12.5
Efficiency (thermal)	---	0.95-0.98	0.89-0.96[c]	0.96-0.97[c]
Environmental Factors				
SO_2, (lb/10^6 Btu)	0.35	0.45	0.35	0.35
NO_x, (lb NO_2/10^6 Btu)	0.40	0.8	0.16	0.16
Particulates, (lb/10^6 Btu)	0.06	0.02	0.02-0.1[d]	0.02-0.1[d]
Solid Waste, ft^3/MW-day	---	25	15	45
S Removal				
Stone	NA		limestone	limestone
Ca/S	NA		~ 15[e]	3.0
Make-up Ca/S	NA		1.0	NA

[a] Basis: 3% sulfur, 90% sulfur removal, 600 MW capacity, 80% load factor.
[b] Equipment modifications are required when converting from gas or coal to low-sulfur oil.
[c] Overall efficiency is dependent on mode of temperature control.
[d] 0.02 figure based on installing electrostatic precipitator (ESP).
0.1 figure based on installing high efficiency cyclone before burners and no ESP.
[e] Ca/S rate dependent on regenerator temperature control scheme.

costs are extremely high with stack gas cleaning, especially on the retrofit case, while operating costs are very near those estimated for oil gasification/desulfurization.

Table 5 lists some advantages of atmospheric-pressure oil gasification over stack gas wet scrubbers.

TABLE 5

ADVANTAGES OVER STACK GAS WET SCRUBBERS

- **Corrosion and fouling problems minimized in SO_2 removal process and boiler (minimum SO_x and V)**
- **No flue gas reheat required**
- **Uses crushed limestone - no limestone pulverizing system needed**
- **Simplified disposal - dry solids and no disposal pond**
- **More compact system**
- **Reduced structural costs**
- **Lower auxiliary power requirement**
- **Reduced energy cost**
- **Improved NO_x control**
- **Potential market for spent CaO**

2.1.4.2 Industrial Boiler Application

Application of the atmospheric-pressure oil gasification system was also considered for industrial boiler operation. Capital and operating costs were projected and compared with alternative sulfur control methods -- low-sulfur oil, fluid bed combustion, and stack gas cleaning. The results are summarized in Tables 6, 7, and 8. Cost and feasibility factors were also identified:

- Load Factor. Decreasing load factor favors clean fuel .
- Space. Decreasing space availability favors clean fuel .

- Reliability. Increasing reliability requirement favors clean fuel.
- Fuel Availability. Decreasing availability of low-sulfur fuel favors use of fluidized bed oil gasification or combustion -- not wet scrubbing.
- Sulfur in Fuel. Increasing sulfur may result in higher increase in cost for wet scrubbing due to disposal pond capacity.
- Nitrogen Oxide. Decreasing NO_x emissions favors oil gasification.
- Particulates. Decreasing particulate emissions favors low-sulfur oil.
- Solids Disposal. Decreasing solids disposal favors low-sulfur oil.
- Efficiency. Increasing efficiency of process steam generation unit favors low-sulfur oil.
- Operating Cost. Minimizing operating cost favors low-sulfur oil of fluidized bed process (difference not considered significant); increased stone requirements or stone cost for fluid bed systems would result in low-sulfur oil being more attractive.
- Capital Cost. Decreasing capital cost favors low-sulfur oil.

Industrial boilers will continue to use clean fuels as long as they are available. If clean fuels are not available, fluidized bed oil gasification or combustion offers economical advantages over competitive systems (e.g. stack gas cleaning). The fluidized bed systems are also able to utilize low-grade fuels (e.g. high-sulfur vacuum residuum) which cannot be easily handled in conventional systems. Thus, industrial applications of fluidized bed oil gasification for industrial fuel gas and/or steam generation offer economical and environmental advantages and should be considered if clean fuels are not available.

2.2 DEMONSTRATION PLANT PROGRAM

2.2.1 Scope

A three-phase demonstration plant program has been conceived:

- Phase I - Preliminary design and cost estimate of a demonstration plant installed on an existing boiler, plus the subsequent preparation of a bid package and a definitive estimate for Phase II work
- Phase II - Detailed design and construction of demonstration oil gasification process
- Phase III - Developmental operation of the gasification process and integrated power plant

Westinghouse, as prime contractor to EPA, will subcontract for the engineering services for Phase I. Phase I has two sequential parts:

- A preliminary design and budget estimate of investment for evaluation of the preliminary design, followed by a decision to stop or proceed.
- Preparation of a bid package and a definitive installed cost estimate.

The bid package is to be adequate for obtaining comparable lump sum bids for the Phase II work. New England Electric System, as the cooperating utility, will supply technical information on the Manchester Street power plant;* provide price and availability information on oil fuels which might be utilized in the plant currently and/or in the future; supply information concerning the load requirements placed on the plant; cooperate in the selection of an engineering firm to prepare the preliminary design and bid package; assist in selecting various options in the

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*Site selected by NEES for demonstration plant in Providence, R.I.

TABLE 6

OIL-FIRED INDUSTRIAL BOILERS WITH SULFUR CONTROL, CAPITAL COST[a] ($/lb steam/hr 370,000 lb steam/hr)

	New Installation			Retrofit
	Boiler[b]	Sulfur Control	Total	Total
Low-Sulfur Oil	2.50	---	2.50	2.80
Fluidized Bed Oil Gasification (once-through)	2.50	5.00[c]	7.50	8.00[c]
Fluidized Bed Combustion[d] (once-through)	---	---	7.50	not applicable
Conventional with Wet Scrubbing	2.50	12.60[e]	15.10[f]	16.70[e]

[a] Suggest use of 0.7 exponent on capacity for scaling costs.

[b] Cost data from note (d) modified to account for increased capacity and cost escalation.

[c] Based on projection for 50 MW Manchester Street Plant. Includes complete limestone handling system, dual fuel capability, fans, engineering, and contingency. Does not include site preparation, escalation, or interest during construction.

[d] "Evaluation of the Fluidized Bed Combustion Process," Ⓦ Report to EPA, Nov. 1971. Does not include site preparation, escalation, or interest during construction.

[e] Based on paper by J. K. Burchard, G. T. Rochelle, W. R. Schofield, and J. O. Smith, Oct. 1972. Includes reheat, solids handling, disposal pond, by-pass, engineering, and contingency. Does not include site preparation, escalation, or interest during construction.

[f] Increase over cost presented in note (d) reference due to additional equipment -- reheat, disposal pond, by-pass, limestone unloading; and different source for wet-scrubbing system cost.

TABLE 7

OIL-FIRED INDUSTRIAL BOILERS WITH SULFUR CONTROL, OPERATING COST

¢/10^6 Btu 370,000 lb/hr, New Installation

	Capital Charges[a]	Fuel Cost	Labor	Sorbent[b]	Power	Solids Disposal[b]	Total
Low-Sulfur Oil	5.10	75	4	---	0.5	---	84.60
Fluidized Bed Oil Gasification Once-through	15.50	45	9	6	1	6	82.50
Fluidized Bed Combustion Once-through	11.50	45	9	6	1	6	78.50
Conventional with Wet Scrubbing	31.20	45	9	3.50	1	13.50	103.20

[a] Assumes 16.7% capital charge, 70% load factor; changes in efficiency will result in < 1¢/10^6 Btu change and will be < 0.4¢/10^6 Btu for the following: low-sulfur oil, 88%; oil gasification, 84%; fluidized bed combustion, 86%; and wet scrubbing, 83%.

[b] \$5/ton for new stone and \$5/ton to dispose of spent stone or slurry; ∿ 2-1/2 Ca/S for fluidized bed systems.

TABLE 8

OIL-FIRED INDUSTRIAL BOILERS WITH SULFUR CONTROL, COST DATA SUMMARY

Boiler Capacity: 100,000 to 500,000 lb steam/hr[a]

	Low-Sulfur Oil	Fluidized Bed Oil Gasification[b] Once-through	Fluidized Bed Combustion[c] Once-through	Conventional Wet Scrubbing
Capital Cost Index	1	3.0	2.3	6[d]
Operating Cost Index	1	0.97	0.93	1.22[c,d]

[a] The relative system costs are assumed to hold over this range of capacity and for either new or retrofit.

[b] Based on cost projection for 50 MW Manchester Street Plant. Includes complete limestone handling system, dual fuel capability, fans, engineering, and contingency. Does not include site preparation, escalation, or interest during construction.

[c] "Evaluation of the Fluidized Bed Combustion Process,' Ⓦ Report to EPA, Nov. 1971. Does not include site preparation, escalation, or interest during construction.

[d] Based on paper by J. K. Burchard, G. T. Rochelle, W. R. Schofield, and J. O. Smith, Oct. 1972. Includes reheat, solids handling, disposal pond, by-pass, engineering, and contingency. Does not include site preparation, escalation, or interest during construction.

design of the system; and evaluate, in cooperation with Westinghouse, Esso (England), and EPA, the effectiveness and economy of the oil gasifier/desulfurizer in power generation and pollution abatement.

The design effort will be based on experimental data from Esso (England) on their 1 MW continuous pilot plant, the conceptual design and assessment by Westinghouse, and operating and environmental constraints established by NEES and local regulations. EPA will provide general guidance and funding for Phase I.

If the preliminary design confirms the effectiveness and economics of the system, a proposal will be prepared for Phases II and III. In Phase II the utility would work closely with an engineering firm in the design and installation of the gasifier/desulfurizer system and share in the costs of the design installation.

In Phase III the plant would be operated to collect basic data on the operation of the gasifier/desulfurizer system, data would be interpreted and analyzed, and an evaluation of the effectiveness (technical and economic) of the process in power generation and pollution abatement would be made.

2.2.2 Location of Utility Partner

Westinghouse, under contract to the Office of Research and Development, EPA, has located a utility partner to carry out a demonstration plant program. New England Power Service Company of New England Electric System (NEES) has agreed to participate in the atmospheric-pressure fluidized bed oil gasification demonstration plant project. The procedure and preliminary design information developed for selecting a utility partner are presented.

2.2.2.1 Utility Presentations

The effort to assemble a project team was initiated by identifying potential utilities and by preparing a utility presentation document -- Appendix E. The technical assessment portion of the overall presentation was prepared to give a review of the oil fuel situation; a general coverage of the atmospheric-pressure oil gasification concepts; the conceptual design evaluation, and the demonstration plant objectives; and coverage of the technical aspects of the program, including an analysis of the Esso (England) data and projected work. A proposed three-phase demonstration plant program is also outlined. A list of utility prospects and the basis for their selection was prepared. A summary of this assessment is presented in Appendix F. Initial utility contacts were initiated on schedule in February 1972, and the first utility presentation was March 7, 1972. Presentations were made to fourteen utilities (Appendix G) who currently operate or plan to operate oil-fired power plants. A general agreement was drafted and presented to each utility at the end of each presentation. This document, presented in Appendix H, served to describe the overall program and indicate the utility participation in each phase of the program. Utility manpower requirements for Phase I (Appendix I) were projected and submitted to utilities expressing interest. Follow-up meetings were held with utilities to obtain their assessment of the process and their possible interest. A list of points which were considered during these meetings was prepared (Appendix J).

2.2.2.2 Utility Response

New England Electric System (NEES) has agreed to proceed with the program. Discussions with NEES are presented in the following section. The Los Angeles Department of Water and Power (LADWP) subsequently expressed its desire to participate in the demonstration plant program. Participation and support in the demonstration plant program with NEES are being pursued with LADWP and other utilities. All of the utilities expressed interest in the concept and considered the process very promising. Numerous reasons were given by those utilities who decided not to participate. The primary reasons appear to be:

- The feeling that funding for such a project should come from Edison Electric Institute (or Electric Power Research Institute).

- Utility involvement in the pursuit of alternative technology or methods for clean power (e.g., stack gas scrubbing, desulfurized oil, methanol), and therefore, not being in a position to take on another development effort
- The need for continuing evaluation of alternative fuel/power conversion systems before making decision
- Shortage of capital.

Several technical concerns were expressed during the discussions with utilities. Primary areas of concern included:

- Hot fuel gas piping valves
- Control and emergency conditions
- Space requirements
- Solid waste disposal
- Modifications and timer required for modifications to boiler
- Plant availability.

These and similar technical concerns will be evaluated in detail during Phase I of the program.

2.2.2.3 New England Electric System

The initial oil gasification system program presentation was made to NEES on July 11, 1972.

NEES expressed considerable interest in the project, and a repeat presentation was made on November 8. Two plants were suggested for the project: Manchester in Providence, R. I. and Salem Harbor 1 in Salem, Mass.

Westinghouse visited the proposed sites and received plant drawings in order to assess space requirements. Space feasibility studies were made which indicated the space was adequate. Economic and design information for each plant was prepared by Westinghouse. A record of these meetings and studies is available elsewhere.[5] NEES selected a 50 MW unit at the Manchester Street Station of Narragansett Electric Co.

for the demonstration plant program. Results of the preliminary studies for the Manchester Street plant are summarized in Appendix L. Alternative stone processing systems to those included in the evaluation are being developed -- both wet and dry processing.

New England Power Service Co. of New England Electric System (NEES) agreed to participate in the atmospheric-pressure fluidized bed oil gasification demonstration plant program on December 11, 1972. The letter of intent from NEES and the letter of acknowledgment from EPA are presented in Appendix K.

The program began with a planning meeting in January 1973. Participants in the program included Ⓦ, NEES, EPA, and Esso (England). An architect-engineer was to be selected to carry out Phase I of the program as a subcontractor to Westinghouse. The nature of the work to be performed is outlined in the invitation to bid package presented in Appendix L.

2.2.3 Conclusions

- System economics indicate atmospheric-pressure fluid bed oil gasification will meet environmental standards and reduce incremental energy costs by up to 50% as compared with stack gas cleaning.
- No problems have been identified which preclude development.
- A utility -- New England Electric System -- is committed to a demonstration plant program.
- The program for a 50 MW demonstration plant is under way. It will require about three years and is estimated to cost $4 to 5 million.

3. OIL GASIFICATION FOR COMBINED CYCLE POWER GENERATION

A pressurized fluid bed oil gasification process has been proposed and a preliminary assessment carried out. Plant performance, capital costs, and energy costs are examined. The ability of the fluid bed process to meet both pollution regulations and gas-turbine requirements is projected on the basis of Esso (England) atmospheric-pressure data. Capital and energy costs for a pressurized fluid bed combined cycle plant process are compared with a conventional oil-fired plant with wet scrubbing, combined cycle plant operation with No. 2 distillate fuel oil, and conventional pressurized oil gasification processes developed by Shell and Texaco. The fluid bed process has the potential to reduce energy costs > 20% below those of a conventional plant with wet scrubbing and > 15% below those of a combined cycle plant using distillate fuel oil.

3.1 INTRODUCTION

The electrical utility demand for residual oil is projected to increase by approximately 250% by 1980.[1,2,3] This rapid growth rate projection results from the fact that alternative fuels -- gas, coal, nuclear -- are not expected to be able to meet the demand. Low-sulfur residual oil to meet federal and local emission regulations will not meet the regulated demand either, nor will desulfurized oil, without an intense effort. All our available oil cannot currently be used. Heavy petroleum fractions, such as 1050°F$^+$ residuum, cannot be used in conventional power plants and can only be desulfurized at excessively high cost.

Thus, gas producers for making clean, low-Btu gas from oil are becoming more attractive to the electrical power industry as the availability of low-sulfur oil and natural gas decreases, the cost of clean fuels increases, and the cost of removing sulfur dioxide from stack gases increases. A gasification system must provide a clean gas, utilize a wide range of fuels, be reliable, and be competitive in energy cost.

Fluidized bed gasification of oil at elevated pressure can produce a clean, low-Btu fuel gas -- 200 to 500 Btu/scf -- for combined cycle power generation. Oil is added to a fluidized bed of limestone or dolomite with sufficient air -- 15 to 25% of stoichiometric -- to maintain the bed at ∿ 1600°F and gasify the oil to produce a fuel gas. The limestone or dolomite removes the sulfur from the fuel gas during the gasification process. The potential for high efficiency, low capital cost, and low emissions makes this system attractive for producing electrical energy from low-cost, relatively abundant, high-sulfur oil.

An evaluation of the pressurized oil gasification system is presented in order to provide a basis for further development. A conceptual design for a system has been developed. Preliminary cost and performance estimates for the pressurized fluid bed gasification process have been developed. These costs are combined with equipment and operating costs for the Westinghouse PACE (Power at Combined Efficiency) plant to obtain total plant and energy costs. The effects of gasification system operating conditions -- e.g. pressure drop and fuel gas temperature -- on plant performance are also evaluated. The pressurized oil gasification combined cycle plant costs are compared with alternative oil-fueled power generation techniques having pollution control. Capital investments and energy costs for a combined cycle plant utilizing pressurized fluid bed oil gasification are compared with those of a conventional oil-fired plant with stack gas cleaning and a combined cycle plant burning No. 2 distillate fuel. The pressurized oil gasification processes proposed by Shell and Texaco are also evaluated and compared with the fluidized bed system. Partial oxidation processes for petroleum are reviewed in Appendix M.

3.2 PROCESS CONCEPTS AND OPTIONS

The pressurized fluid bed oil gasification process shown in Figure 4 consists of a fluidized bed gasifier/desulfurizer vessel, a limestone/dolomite regeneration section, and a sulfur recovery section. The phenomena taking place in the pressurized fluid bed gasifier are

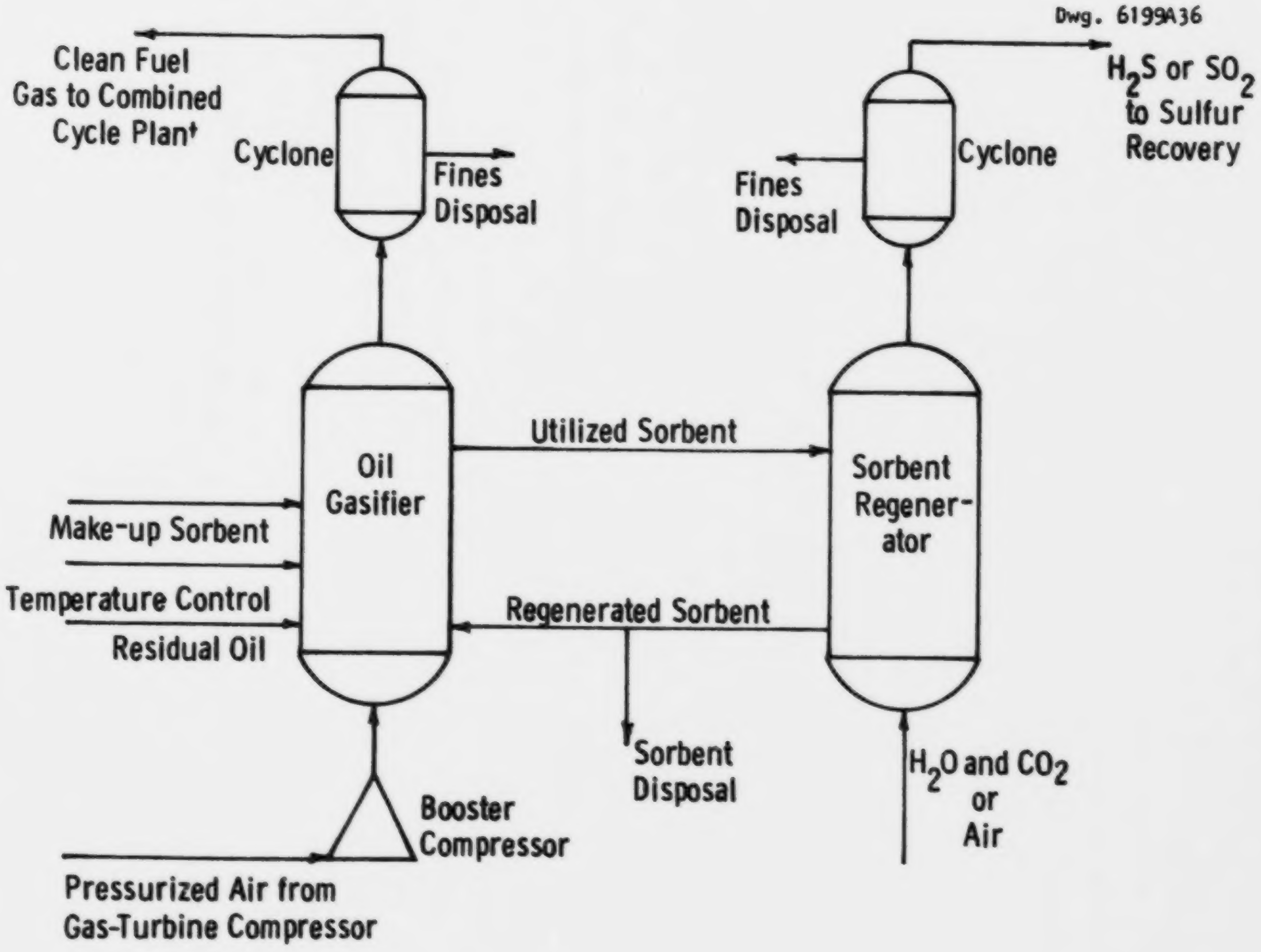

Figure 4 —Regenerative Pressurized Oil Gasification Process

essentially the same as have been described for the atmospheric-pressure fluid bed case.[6,7,8] Residual oil is injected into the gasifier vessel, an air-fluidized bed of lime at 1600°F operated with substoichiometric air (14 to 25% of stoichiometric). This yields (by cracking, partial combustion, and H_2S absorption by the lime or dolomite), a hot, low-Btu, fuel gas and sulfided lime or dolomite. The fuel gas is transported to the combined cycle plant where combustion is completed, and the sulfided lime or dolomite is sent to the regenerator. The regenerator may be designed to produce an H_2S or an SO_2-rich gas and a regenerated sorbent with a decreased activity (when compared to that of fresh sorbent). The SO_2 or H_2S stream is transported to a sulfur recovery system, and the regenerated sorbent is returned to the gasifier along with a nearly stoichiometric amount of fresh make-up sorbent.

A variety of temperature control schemes may be useful in controlling the temperatures of the gasifier and the regenerator. The temperature control scheme used in the gasifier must conserve the energy of the fuel in a form usable in the boiler. For this reason water injection, steam injection, and stack gas recycle are considered for temperature control of the gasifier by the storage of sensible heat. Both limestone and dolomite are considered as sulfur absorbents in the pressurized fluid bed case. The major process options which have been examined with respect to cost and performance follow.

3.2.1 Gasifier Air/Fuel Ratio

Air/fuel ratios of 14% and 25% of stoichiometric giving fuel gas low heating values (hot) of about 500 Btu/scf and 280 Btu/scf, respectively, have been considered. An air/fuel ratio of 14% of stoichiometric is assumed to be the minimum air/fuel ratio at which a gasifier temperature of 1600°F can be maintained, while 25% of stoichiometric is assumed to be a conservatively high air/fuel ratio, according to the experience gained from the Esso (England) atmospheric-pressure operations. The physical feasibility of operating at an air/fuel ratio as low as 14% of stoichiometric without excessive carbon deposition in the gasifier must be demonstrated.

3.2.2 Limestone/Dolomite Regeneration Method

The regeneration of calcium sulfide and the production of a high-sulfur gas for sulfur recovery can be achieved by either of two processes, as follow:

3.2.2.1 Regeneration with CO_2 and H_2O to Produce H_2S

With this process the reaction

$$CaS + H_2O + CO_2 \rightleftarrows CaCO_3 + H_2S$$

is utilized to produce calcium carbonate with H_2S-rich gas. The reaction is favored by high pressure and would be carried out at a temperature of about 1100°F, with H_2S concentrations of about 9% (vol) being projected. Two sources of CO_2 for the regeneration scheme have been considered:

- By scrubbing flue gas from the combined cycle plant
- By scrubbing a gas stream produced by combustion of a portion of the fuel gas.

3.2.2.2 Regeneration with Air to Produce SO_2

The reaction of oxygen with calcium sulfide,

$$CaS + 3/2\ O_2 \rightleftarrows CaO + SO_2 \quad ,$$

is favored by low pressure and high temperature (∿ 2100°F). Although this regeneration scheme would yield an SO_2 concentration of only about 2% (vol) at pressures of 10 to 15 atm, it is considered because of its apparent simplicity.

3.2.3 Once-through System

If a high sorbent make-up rate is required for the regenerative processes and if high sorbent utilization can be achieved in the gasifier, a once-through system may be attractive. Once-through operation would require conversion of calcium sulfide to calcium sulfate before disposal of the sorbent. The reaction

$$CaS + 2\ O_2 \rightleftarrows CaSO_4$$

could be applied for this purpose and would be carried out at 1400°F to 1700°F. Calcium utilization in a once-through process is expected to be 35% higher.

3.2.4 Other Options

Other process options examined are the pressure drop required across the gas-turbine combustor and the option of cooling the gas produced by fluidized bed oil gasification before it is combusted in the combined cycle plant.

3.3 PROCESS SPECIFICATIONS AND DESIGN BASIS

Table 9 lists the factors specified for the conceptual design of the pressurized fluid bed oil gasification process. The combined cycle power generating plant is a Ⓦ PACE plant consisting of two Ⓦ-501B gas turbines and a single 28.5 in. steam turbine. The pressure of the fuel gas to the gas turbine is 215 psia. Designs and energy costs are based on an operation using limestone for desulfurization. Sulfur removal of 90 to 95% from a 3 wt % sulfur residual oil is specified for the fluid bed process, with the specifications for vessel temperatures, limestone utilizations, and the limestone make-up rate based both on thermodynamic information and on the atmospheric-pressure data of Esso (England). Control of the vessel temperature is assumed to be carried out easily by means of any of the methods suggested in the table, based on atmospheric information. A plant heat rate of 9000 Btu/kWh was assumed for the purpose of finding the approximate fuel consumption of a PACE plant.

3.4 MATERIAL AND ENERGY BALANCES

Material and energy balance information for the pressurized fluid bed oil gasification process was based on Esso (England) atmospheric-pressure data. Performance and vessel sizes were modified for effects of pressure. Regeneration system designs were based largely on

TABLE 9

SPECIFICATIONS FOR FLUID BED OPERATION

Item	Specification
Plant Electrical Capacity	250 MW (PACE Plant)
Capacity Factor	70%
Turndown Ratio	4/1
Number of Gasifier Modules	2
Modes of Operation	regenerative or once-through
Gasifier Pressure	∿ 15 atm
Pressure of Gas Turbine	11 atm (165 psia to turbine)
Sulfur Removal	90-95%
Residual Oil	3 wt % sulfur
Gasifier Temperature	1600°F
Regenerator Temperature	1100°F[a]
Sulfate Generator Temperature	1700°F with once-through option
Air/Fuel Ratios	14 (minimum) & 25% stoichiometric
Lime Particle Diameters	500-2000 μ average diameter
Regenerative Lime Utilization	10%
Once-through Lime Utilization	35%
Limestone Make-up Rate	1 mole CaO/mole sulfur fed
Gasifier Temperature Control	stack gas recycle, steam or water injection
Regenerator Temperature Control	lime circulation rate; water injection
Sulfate Generator Temperature Control	excess air circulation (∿ 200% excess)
Plant Heat Rate	9000 Btu/kWh (assumed for purposes of material balances)

[a] With CO_2/H_2O regeneration, 2100°F with air regeneration.

thermodynamic behavior. The general behavior of the fluidized bed gasifier for sulfur removal, vanadium removal, and carbon deposition is assumed to be independent of pressure, once the gasifier bed diameter and bed depth has been scaled for the affect of pressure. Details are presented in Appendix N.

Figure 5 shows a simple block flow diagram for the pressurized fluid bed oil gasification process. Block diagrams of this type have been utilized to analyze the performance of the gasification process and the performance of the complete power plant.

Inputs to the fluidized bed process are residual oil; air; water and steam; limestone (or dolomite); booster compressor power; auxiliary power for pumping oil, water, and solids circulation; and gas compression in the regeneration section of the gasification system. Output is hot fuel gas for the combined cycle plant; sulfur; and energy losses in the form of sensible heat of the spent limestone, heat losses, carbon deposition losses, and sensible heat of flue gas from the regeneration and sulfur recovery sections. The thermal efficiency of the fluidized bed gasification system will depend slightly on the air/fuel ratio and the regeneration method used but will be about 90 to 95% for all the options considered.

Fuel compositions and heat values are shown in Table 10 for the oil gasification process. The product gas compositions for the fluidized bed process at air/fuel ratios of 14% and 25% of stoichiometric have been estimated from atmospheric-pressure data. Projected plant performance has been based on these product compositions.

3.5 CAPITAL INVESTMENT EVALUATION

Figure 6 is a flow diagram for the pressurized fluid bed oil gasification process with limestone regeneration by reaction with CO_2 and H_2O to form $CaCO_3$ and H_2S. The source of CO_2 is flue gas from the combined cycle plant. The alternative CO_2 source, combustion of a

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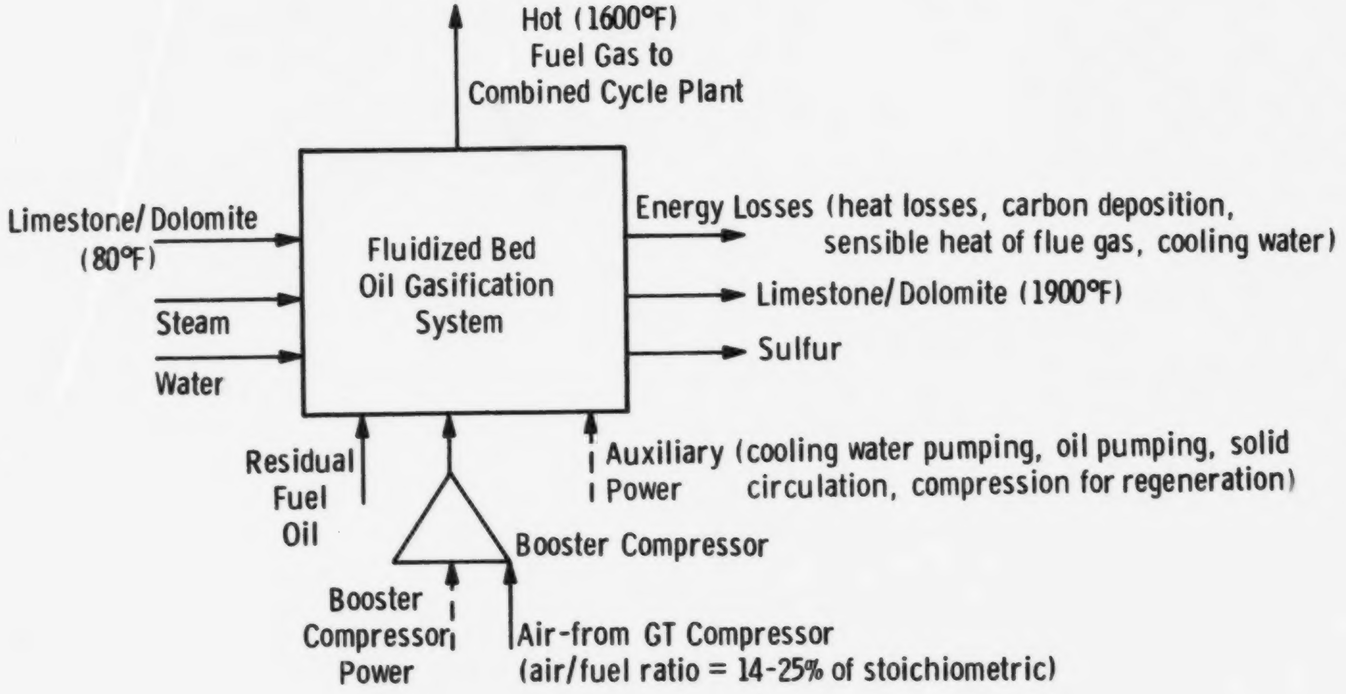
Hot (1600°F)
Fuel Gas to
Combined Cycle Plant
Limestone/Dolomite
(80°F)
Steam
Water
Fluidized Bed
Oil Gasification
System
Energy Losses (heat losses, carbon deposition, sensible heat of flue gas, cooling water)
Limestone/Dolomite (1900°F)
Sulfur
Residual
Fuel
Oil
Auxiliary (cooling water pumping, oil pumping, solid
Power circulation, compression for regeneration)
Booster Compressor
Booster
Compressor
Power
Air-from GT Compressor
(air/fuel ratio = 14-25% of stoichiometric)

Figure 5—Energy Balance for the Fluidized Bed Oil Gasification Process

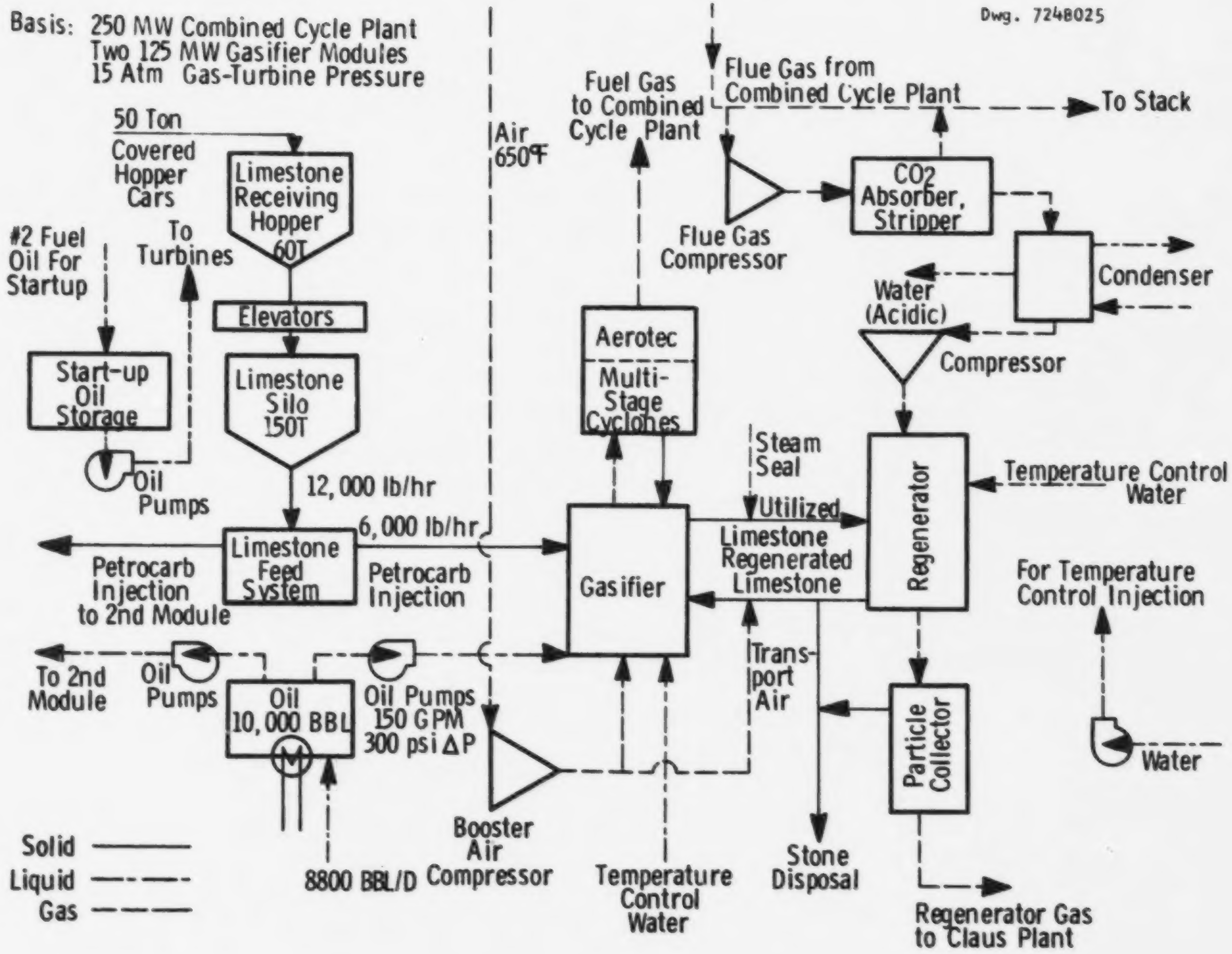

Figure 6—Pressurized Oil Gasification Plant Flow Diagram

portion of fuel gas, has been eliminated on the basis of the comparative economics of the two options. The equipment shown in Figure 6 consists of fuel oil and limestone handling equipment, air booster compressor, gasifier vessels (two modules) with multistage particle collection, and a CO_2/H_2O regeneration system. Gasifier design is considered in Appendix O.

TABLE 10

GASIFICATION PRODUCT COMPOSITIONS

	Fluidized Bed 14% A/F	Process (Vol %) 25% A/F
N_2	50.74	47.74
H_2	0.82	2.64
CO	13.40	7.97
CO_2	6.70	7.97
H_2O	0.00	20.56
CH_4	9.43	6.55
C_2H_4	18.86	6.55
H_2S	0.05	0.02
Low Heating Value (hot), Btu/scf	∿ 500	∿ 280

The process has been separated into four component systems -- the gasification system, the CO_2/H_2O regeneration system, the booster compressor, and the Claus plant. These four component costs are shown in Figure 7 in units of $/MM Btu/hr of oil feed (HHV), as a function of the pressure at the gasifier product gas outlet, with the air/fuel ratio as a parameter. Two cases of pressure drops required in the gas-turbine combustor and the fuel distribution equipment are considered in the figure -- 8 psi ΔP and 53 psi ΔP. The pressure drop across the gasifier and particle control equipment is assumed to be 9% of the gasifier pressure.

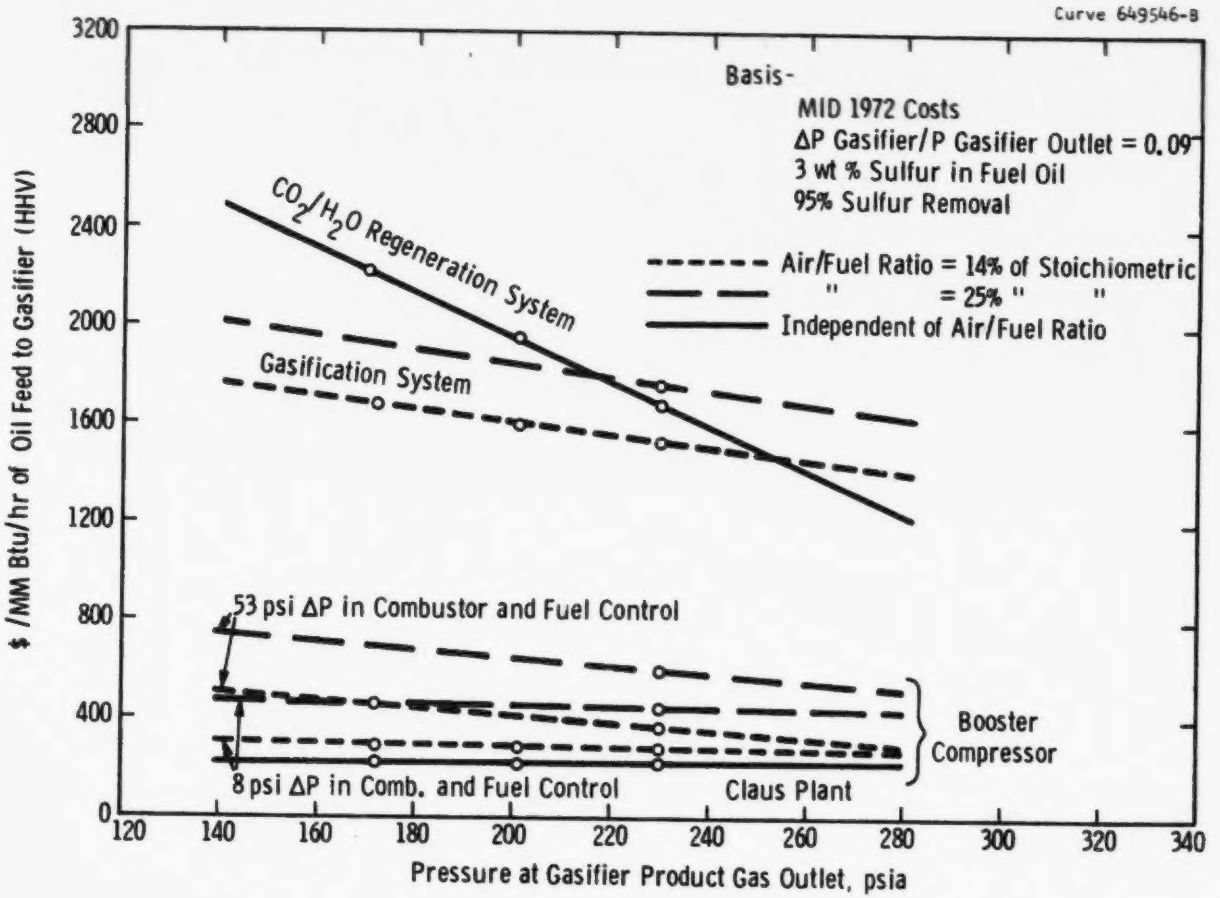

Figure 7—Fluidized Bed Oil Gasification Capital Investment

Figure 7 leads to the following conclusions:

- Reducing the air/fuel ratio from 25% to 14% of stoichiometric reduces the capital investment by about 10%, in units of \$/MM Btu/hr.
- For the case of a 10-atmosphere gas turbine, increasing the combustor pressure drop from 8 psi to 53 psi reduces the process cost by about 10% because of the highly pressure-sensitive nature of the CO_2/H_2O regeneration scheme.
- A once-through scheme would reduce the cost of the process by more than 40%, neglecting the slight cost increase due to a sulfate generating vessel.
- The cost of the fluidized bed process with an air-blown regenerator would be somewhere between the cost for the CO_2/H_2O regeneration system and the once-through gasification system.

The true capital investment for pressurized fluid bed oil gasification will depend on the plant performance as a function of the design options -- air/fuel ratio, combustor pressure drop, and regeneration method. The process cost estimate is discussed in Appendix P.

3.6 PERFORMANCE

Cycle studies based on estimated material and energy balances and approximate gas compositions (Table 10) have led to estimates of the plant heat rate, the power plant capital investment, and energy costs, using pressurized oil gasification with combined cycle power generation. Detailed results are shown in Appendix Q. The following conclusions concerning the combustor pressure drop have been deduced:

- With the pressurized fluid bed oil gasification process, an increase in the required combustor pressure drop of 10 psi increases the plant heat rate by only 10 Btu/kWh with operation at an air/fuel ratio of 14% of stoichiometric, and by 30 Btu/kWh at 25%.

- With the pressurized fluid bed oil gasification process, an increase in the required combustor pressure drop of 10 psi reduces the plant power by about 0.5 MW with operation at 14% of stoichiometric, and about 1.0 MW at an air/fuel ratio of 25% of stoichiometric.

From these points it is concluded that the combustor pressure drop is a fairly insensitive parameter with respect to plant performance and should be determined by the requirements for fuel distribution and control.

Table 11 summarizes the performance, capital cost, and energy cost of the fluidized bed process with a combustor pressure drop of 50 psi for the two air/fuel ratios, for regenerative operation by CO_2/H_2O, and for once-through operation. The figures shown in the table refer to the case in which the 1600°F fuel gas from the fluid bed process is not cooled before combustion. Table 11 indicates that the plant heat rate with the fluid bed process increases with increasing air/fuel ratio, while the heat rate is comparable for the regenerative and once-through operations. The plant heat rate is less than 10,000 Btu/kWh even for the high air/fuel ratio case of 25% of stoichiometric. Plant power is comparable for all of the cases shown. Capital investment is based on PACE plant cost information and the cost estimates shown in Figure 7. The cost basis is listed in Table 11. Capital cost is reduced for the fluid bed process with once-through operation because of the great expense involved in limestone regeneration by the CO_2/H_2O method. The change in plant capital cost in going from 14% air/fuel ratio to an air/fuel ratio of 25% of stoichiometric is slight.

Energy cost assumptions are listed in Table 11. The 14% air/fuel ratio operation of the fluid bed process is more economical than the 25% air/fuel ratio, while the energy cost for once-through operation is also less than that for regenerative operation. The relationship between energy cost of regenerative and once-through operation is dependent on the cost of limestone disposal; but even if the limestone

TABLE 11

PROCESS PERFORMANCE[a]

	Fluid Bed Process CO_2/H_2O Limestone Regeneration		Fluid Bed Process Once-through Operation	
	14% A/F	25% A/F	14% A/F	25% A/F
Plant Heat Rate (HHV), Btu/kWh	9007	9828	8906	9716
Plant Power, MW	266.0	276.0	269.0	279.0
Plant Capital Cost, $/kw	179.5	182.1	158.8	159.5
Total Energy Cost, mills/kWh	9.43	9.93	8.96	9.42
Break-even Distillate Cost, ¢/MM Btu	65.2	71.1	59.7	65.1

[a] Conditions Assumed

- 50 psi combustor pressure drop
- 1600°F fuel gas temperature from fluid bed process with no cooling
- Fixed charges at 15%
- Residual oil at 45¢/MM Btu
- Limestone at $6/ton
- 5% contingency; 4% escalation; 8% interest during construction; 2 years construction time; 2% A&E; 70% capacity factor
- No credit for sulfur recovered.

cost should increase to $17/ton the once-through and regenerative system energy costs would be about equal. Calcium utilization greater than 35% would reduce the once-through energy cost further.

Break-even costs for No. 2 distillate fuel oil is given in ¢/MM Btu and represent the price at which No. 2 distillate must be available for the PACE plant to operate at the same energy cost as the residual oil gasification. With No. 2 distillate prices ranging from 85¢ to 95¢/MM Btu the fluid bed process is expected to be more economical than PACE plant operation with No. 2 distillate.

The performance of the fluid bed oil gasification process has been examined with respect to three methods of cooling the 1600°F product gas: cooling the gas and discarding the sensible heat; cooling the gas and recovering sensible energy at an efficiency of 25%; and cooling the fuel gas by direct cooling with injection of 77°F water. The following general conclusions have been derived from this examination:

- For both regenerative and once-through operations with the fluid bed process, cooling the 1600°F fuel gas to 400°F and discarding the energy will increase the plant heat rate by about 1600 Btu/kWh, increase the energy cost by about 0.9 mills/kWh, and yield nearly the same plant power, with an air/fuel ratio of 25% of stoichiometric. The corresponding increases with an air/fuel ratio of 14% are about 800 Btu/kWh and 0.45 mills/kWh.
- For both regenerative and once-through operations with the fluid bed process, injection of water to cool the 1600°F fuel gas to 400°F increases the plant heat rate by about 3,500 Btu/kWh, increases the plant power by about 140 MW, and increases the energy cost by about 1.0 mills/kWh, with an air/fuel ratio of 25%. The corresponding increases with a 14% air/fuel ratio are 1500 Btu/kWh, 50 MW, and 0.49 mills/kWh.

- For both regenerative and once-through operations with the fluid bed process, cooling the 1600°F fuel gas to 400°F with a waste heat boiler and recovering the energy with an efficiency of 25% will increase the plant heat rate by about 550 Btu/kWh, and increase the power plant by about 35 MW, with a 25% air/fuel ratio. The corresponding increases with a 14% air/fuel ratio are 300 Btu/kWh and 15 MW. The energy cost increase is slight if a reasonable cost is assumed for a conventional waste heat boiler system and if it is assumed that cooling of the hot fuel gas is feasible at air/fuel ratios as low as 25% of stoichiometric -- about 0.10 mills/kWh at 14% air/fuel ratio and 0.16 mills/kWh at 25% air/fuel ratio. In reality, the hot fuel gas cannot be cooled in a conventional waste heat boiler system because of the presence of tars (∿ 5 to 10 wt %) which will condense on all cooling surfaces. Cooling of the gas will require either operating at much higher air/fuel ratios (∿ 40% of stoichiometric) or quenching and scrubbing the hot gas to remove tars. The first option would increase the gasifier capital about $10/kw (not including the cost of a waste heat boiler system and soot removal equipment) while reducing the plant efficiency. The second option would result in inefficient operation with high energy losses, as with cooling of the gas by water injection. Both options seem unattractive for power generation, though further analysis is required to develop quantitative conclusions.

3.7 COMPARISONS WITH ALTERNATIVE POWER GENERATION SYSTEMS

Tables 12 and 13 compare capital investments and energy costs of pressurized fluid bed oil gasification with alternative oil-fueled, pollution controlled, power generation systems. Capital and energy costs for a conventional oil-fuel power plant utilizing limestone wet scrubbing for pollution control and a PACE plant fueled with No. 2 distillate, are compared with capital and energy costs for a PACE plant with pressurized

TABLE 12

CAPITAL INVESTMENT COMPARISONS

	Conventional Oil-Fired Power Plant with Scrubber	PACE Plant with No. 2 Distillate Fuel	PACE Plant with Residual Oil Gasification[a]
Total Capital Cost ($/kw)[b]	323.28	137.80	182.1
Assumptions			
Construction Time (years)	4.5	2.0	2.0
Sulfur Removal Equipment ($/kw)	50.0	--	44.3
Plant Capacity (MW)	635	269	269

[a] CO_2/H_2O regeneration of limestone and air/fuel ratio of 25% of stoichiometric -- no cooling.

[b] 5% contingency; 4% escalation per year; 8% interest during construction; 2% A&E 70% capacity factor.

TABLE 13

ENERGY COST COMPARISONS -- mills/kWh[a]

	Conventional Oil-Fired Power Plant with Scrubber	PACE Plant with No. 2 Distillate Fuel	PACE Plant with Residual Oil Gasification
Fixed Charges	7.91	3.37	4.45
Fuel	4.11	7.80	4.42
Limestone	0.12	--	0.13
Operating and Maintenance	0.91	0.52	0.93
TOTAL (mills/kWh)	13.05	11.69	9.93

[a] Assumptions

- Fixed charges at 15%; 70% capacity factor
- 3 wt % sulfur residual oil at 45¢/MM Btu
- Limestone at $6/ton
- No. 2 distillate at 90¢/MM Btu
- No credit for sulfur recovered

fluid bed oil gasification. The fluid bed process is carried out at a 25% air/fuel ratio with limestone regeneration by the CO_2/H_2O method, representing the highest capital cost and lowest efficiency of the fluid bed cases considered. Due to the low cost of the PACE plant package, the PACE plant with fluid bed oil gasification is about $140/kw cheaper than a conventional power plant. Energy cost of the PACE plant with fluid bed oil gasification, assuming 45¢/MM Btu residual fuel and 90¢/MM Btu distillate fuel, is more than 3 mills/kWh less than the energy cost of the conventional power plant and about 1.7 mills/kWh less than a PACE plant fueled with No. 2 distillate.

3.8 SHELL AND TEXACO PRESSURIZED OIL GASIFICATION PROCESSES

A pressurized oil gasification process of the type which has been operated for gas manufacture by Shell[9] and Texaco[10] has been proposed for the production of a low-Btu gas for combined cycle power generation. A brief description and assessment of the processes are given to indicate the performance of a conventional oil gasification technique. The Shell and Texaco processes generate a low-temperature (100 to 250°F), clean fuel gas having a low heating value of about 120 Btu/scf. Steam generated by waste heat boilers in the gasification process is also provided for the combined cycle plant.

A flow diagram for the Shell and Texaco processes is shown in Figure 8. The Shell and Texaco processes consist of an air-blown oil gasification vessel (partial oxidation reactor) operated at a temperature of about 2500°F with an air/fuel ratio of about 45% of stoichiometric. The hot gas is cooled in waste heat boilers, producing saturated steam, prior to purification of the gas. The gas-purification process consists of a carbon (soot) recovery step and a sulfur-removal section. Recovery soot (∿ 3 wt % of the fuel oil feed rate) is recycled to the gasifier vessel, and H_2S produced in the sulfur removal section is sent to a

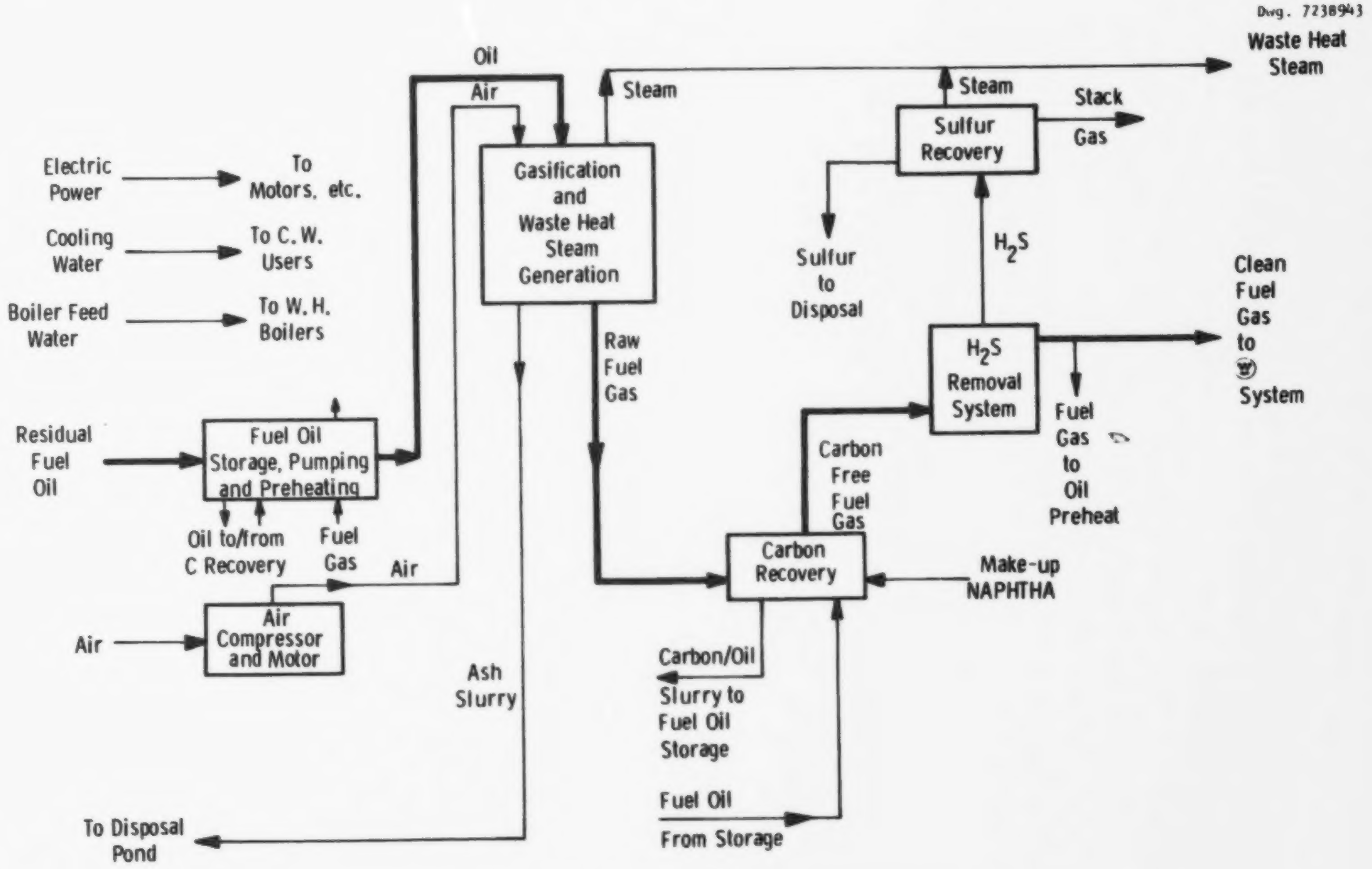

Figure 8—Pressurized Oil Gasification for Gas Turbine Fuel - Shell and Texaco Processes

sulfur recovery plant to recover elemental sulfur. The cool fuel gas (100 to 250°F) and the saturated steam produced by the process are sent to a combined gas-turbine/steam-turbine power generation system.

The composition of the cool fuel gas is shown in Table 14. The gas has been dried and desulfurized to about 5 ppm H_2S. On the basis of this fuel gas composition and an assumed gasification process thermal efficiency of 87%, a plant heat rate of about 11,000 Btu/kWh is projected for utilization with present combined cycle technology. Thermal efficiencies as low as 80% have been projected for these processes.

With a capital cost of about $91/kw assumed for the Shell/Texaco processes and by application of the same cost basis used in Table 13, the energy cost breakdown in Table 15 is calculated.

TABLE 14

SHELL/TEXACO PROCESSES PRODUCT COMPOSITION

Product	Volume %
N_2	60.79
H_2	14.57
CO	22.92
CO_2	1.37
H_2O	0.00
CH_4	0.35
C_2H_4	0.00
H_2S	5 ppm
Low Heating Value, Btu/scf	117

TABLE 15

SHELL/TEXACO ENERGY COST

Item	Mills/kWh
Fixed Charges	6.21
Fuel	4.95
Chemicals and Catalysts	0.06
Operating and Maintenance	1.75
TOTAL	12.97

This analysis is based on information provided by Shell and Texaco about their processes (Appendix R and references).

3.9 CONCLUSIONS

- Pressurized fluid bed oil gasification with combined cycle power generation will deliver energy at costs > 20% less than power generation with clean fuels.
- Combined cycle power generation utilizing the Shell and Texaco processes, as they have been conceived in this study, are not competitive with alternative power generation processes. The processes might be altered, however, to be more compatible with power generation requirements and to improve system efficiency.
- A program should be initiated to develop pressurized fluidized bed oil gasification for combined cycle power generation.

4. SUPPORT STUDIES

4.1 INTRODUCTION

Hydrogen sulfide produced during the gasification of oil reacts with lime to form calcium sulfide:

$$H_2S + CaO \rightarrow CaS + H_2O \quad .$$

Fluidized bed desulfurization of oil is currently under study at Esso (England), with a view to determining the probable operating conditions and behavior of a commercial system in which the lime sorbent is regenerated.

The variety of process options under consideration makes it desirable that the above reaction be well characterized and that alternative schemes for ultimate disposal of the fuel sulfur be assessed.

Laboratory studies on the reaction by Squires and his school have demonstrated the utility of thermogravimetric analysis in examining the behavior of this reaction. As part of a laboratory program to investigate the reactions between calcium carbonate, or lime, and the sulfur gases obtained by gasification or combustion of fossil fuels, Westinghouse has constructed a thermogravimetric analysis system with the capability of studying these reactions over a wide range of temperatures, pressures, and gas compositions. (The system is described in detail in Volume I, Section 3.2.)

The ultimate aim of the thermogravimetric studies is to obtain data which can be used to predict the fluidized bed performance of the reactions, for specific limestones. The intermediate objectives are:

1. To establish which reactions occur
2. To determine the kinetics of the reactions and the stone utilization achievable

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3. To determine the effect of possible process operating conditions on (2)
4. To probe reaction mechanisms
5. To assess the influence of side reactions
6. To recommend optimal operating conditions.

In the work reported here, the sulfidation of limestones and dolomites at atmospheric pressure and 10 atmospheres is described in 4.2. The alternative to sorbent regeneration is explored in section 4.3 where studies on the oxidation of calcium sulfide, as produced by a once-through process, are reported.

4.2 SULFIDATION OF CALCINED LIMESTONE AND DOLOMITE

This study was undertaken

- To extend Westinghouse knowledge of sulfidation of limestone and dolomites to pressurized conditions
- To examine the sulfidation characteristics of particular samples of limestone and dolomite
- To provide substrates for other reactions of interest (oxidation of calcium sulfide; regeneration of calcium carbonate).

In the course of this study, some features of the sulfidation reaction have been explored. While the study is as yet incomplete, the principal characteristics of the reaction have been demonstrated. The results confirm the general conclusions of Pell[1] and extend knowledge of the system to reaction under pressure and to a wider range of substrates.

4.2.1 Conclusions from This Study

- In the TG apparatus at 1600°F, the sulfidation of limestones and dolomites is controlled by convective mass transfer from the gas stream to the solid.
- As reaction proceeds, diffusion of hydrogen sulfide through a product layer becomes a rate-limiting step.
- Sulfidation at pressure (10 atmospheres) shows little or no increase over the rate at atmospheric pressure.
- The rate of reaction observed is rapid (virtually all the calcium may be sulfided), and should yield high-sulfur retention in fluidized beds of limestone and dolomite, in accord with the experience of other workers.

4.2.2 Experimental Methods

The apparatus and equipment is described in 4.2.13.

Hydrogen sulfide was introduced as 15% H_2S in nitrogen, obtained from Air Products and Chemicals, Inc. It was metered with reference to nitrogen flow as calibrated on a Matheson flowmeter calibrator and determined iodometrically. Some sulfur deposition on the quartz reactor always occurred, even when hydrogen was present to suppress dissociation; it was probably the product of heterogeneous reaction on the quartz reactor surfaces.

Stone sorbents were used as received; sieve fractions were obtained by crushing lightly and sieving on stainless steel. The list of sorbents used is appended in 4.2.11.

Calcium was determined by precipitation as the oxalate and ignition to the oxide.[2] Later analyses were carried out according to the similar procedure described in "Methods of Analyzing and Testing Coal and Coke."[3]

The number of runs carried out was curtailed for two reasons:

- Equipment reliability
- Pressure of demand for runs on other reactions in the $CaCO_3/S$ system.

4.2.3 Equipment Problems

Pressurized runs with hydrogen sulfide led to corrosion of the balance mechanism. This was accompanied by fouling of the pressure relief valve. The passage of hydrogen sulfide through the pressure relief valve caused deposition of sulfur on the valve seat, and consequently the valve tended to stick and create a cycle in which the exit was sealed off. The continuous flow of $2\ell.min^{-1}$ of hydrogen sulfide containing reactant gas mixed with the purge gas and corroded the galvanometer coils. Two balance movements were lost in this fashion.

In an attempt to correct this fault, a trap containing a high surface area zinc oxide, heated to 572°F, was installed in the exit line. It has so far prevented any deposition of sulfur in the pressure relief valve. The useful life of a balance under these conditions cannot yet be predicted. A new design for the corrosion prevention system is needed.

In the initial phase of our experimental program, a gas chromatograph was attached to the effluent line; however, after some experimentation, its use was discontinued. The problems of operating the TG apparatus at pressure required undivided attention.

Sulfidation runs can be divided into three phases:

4.2.4	Early Experimental Runs Establishing Procedures	Phase 1
4.2.5	Calcined Dolomite at Atmospheric Pressure	
4.2.6	Calcined Limestone 1359 at Pressure	Phase 2
4.2.7	Runs at Pressure on Sulfidation of Dolomite	
4.2.8	Atmospheric Sulfidation of Limestones Submitted by NEES	Phase 3 .

4.2.4 Atmospheric Sulfidation of Tymochtee Dolomite

Dolomite apparently calcines at a lower temperature in hydrogen sulfide than in air when heated at a constant rate (15 $K.min^{-1}$); in 15% H_2S in nitrogen, the calcination temperatures for otherwise similar conditions are depressed by 210°F. After a loss in weight equivalent to decomposition of the magnesium fraction, the reaction decelerates through a temperature of 1292°F. Further heating causes acceleration of the reaction and then a change from weight loss to weight gain.

This experiment is explained by the following reaction sequence, which is not, however, unique; because of the large number of experiments required to verify the process during a nonisothermal sulfidation, the system was not considered worthy of attention for its own sake.

- $MgCO_3$ in dolomite calcines. The presence of H_2S apparently catalyzes this step. (The presence of different gases is known to catalyze calcium carbonate decomposition.)[4]
- The calcite fraction reacts with hydrogen sulfide and, because of the high concentration of hydrogen sulfide, the reaction, as Pell[1] found, is self-poisoning at 1292°F.
- Above 1292°F, reaction of H_2S with $CaCO_3$ again proceeds, slows down as the calcite is exhausted and the MgO slowly sulfides, yielding a weight gain.
- On standing in nitrogen, traces of moisture hydrolyze the magnesium sulfide which reforms the magnesium oxide.

4.2.5 Further Isothermal Experiment Results

- After calcination, dolomite sulfides in two phases. The rapid sulfidation corresponds to stoichiometric formation of calcium sulfide. The second reaction was assumed to be slow formation of magnesium sulfide. Polysulfide formation was not considered likely, since the phenomenon did not occur in reactions with limestone. Ruth[5] has since reported that x-ray analysis shows the formation of magnesium sulfide.

- The magnesium sulfide is unstable even to residual traces of moisture in the tank nitrogen. Introduction of water vapor leads to extremely rapid hydrolysis.
- Sulfidation is extremely rapid for 500 μ particles of calcined dolomite. In 15% hydrogen sulfide at 1552°F, reaction was 80% complete within 40 seconds. This rate is in excellent agreement with the model Pell[1] derived for the true reaction rate of hydrogen sulfide with finely powdered calcined dolomite (Figure 9).

4.2.6 Calcined Limestone 1359

Limestone 1359 was calcined at a constant heating rate and then sulfided at temperatures of 1600°F and 1380°F.

Since these were the first experiments carried out at pressure on this apparatus, the experiments were largely directed at establishing operability of the system.

In the first experiments, TG 63 and TG 64, calcination was apparently delayed by pressurization. Sulfidation was remarkably slow by comparison with atmospheric sulfidation of dolomite. In TG 67, 68, and 69 (Figure 10), the gas-flow rate over the sample was varied and the course of reaction was found dependent on flow. A series of experiments was therefore conducted on limestone calcination under pressure.

The results of the experiments on calcination at pressure showed that gas-volume flow rates of $2\ell.\text{min}^{-1}$ were needed to yield, at 10 atmospheres, kinetics close to those observed at 1 atmosphere pressure.

Three further experiments on sulfidation at pressure showed that the sulfidation rate of calcined limestone at 10 atmospheres was comparable to the sulfidation rate of calcined dolomite at atmospheric

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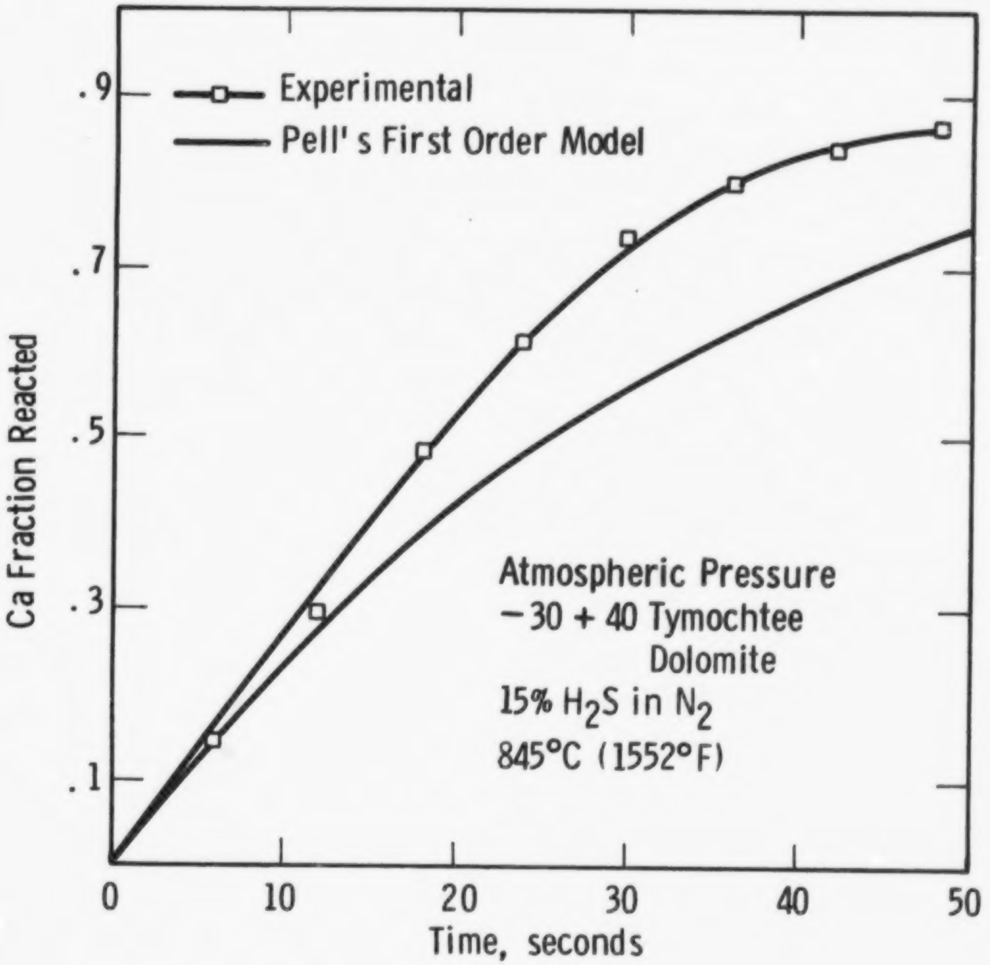

Figure 9—Sulfidation of Calcined Dolomite

pressure. Complete sulfidation required about forty minutes for 500 μ-diameter particles at 0.24% hydrogen sulfide in nitrogen. Formerly the contracting volume expression

$$\frac{1-(1-\alpha)^{1-n}}{1-n} = kt \; ; \quad n = 2/3$$

adequately described the reaction. The similarity in rates, however, between a dolomite substrate at atmospheric pressure and a limestone at 10 atmospheres, and the sensitivity of reaction rate to gas-flow conditions, preclude interpretation of the course of reaction as arising from a depletion of reactant, although the results may be so described empirically.

4.2.7 Sulfidation at Pressure on Dolomite

Sulfidation runs at pressure were carried out on

- 5/6 mesh Tymochtee dolomite (half-calcined)
- 16/18 mesh Glasshouse dolomite (fully calcined)
- 16/18 mesh Tymochtee dolomite (fully calcined).

The result of sulfiding fully calcined dolomite at 10 atmospheres pressure is shown in Figure 11. By comparison with atmospheric-pressure data, the ratio of the initial fastest rates was 1.22 (in agreement with the predictions of a convective mass transfer mechanism).

A slower mode of sulfidation becomes apparent after the calcium content of the stone has been sulfided. The weight gain observed is reversible; when nitrogen (undried) is passed over the sample, it slowly loses weight. This additional weight gain may be attributed to magnesium sulfide formation and to polysulfides of calcium. Ruth[5] has confirmed the presence of magnesium sulfide in sulfided dolomite. Polysulfide formation is unlikely, since limestone 1359 showed no evidence of additional reaction once the stoichiometric formation of calcium sulfide had occurred.

The effect of lowering the temperature to 1450°F is shown in Figure 11. Completion of the series with tests at 1750°F and 1850°F were prevented by an electrical fault. Nonetheless, it can be seen that the reduction in rate on lowering the temperature by 150°F is slight.

4.2.7.1 Sulfidation of Large Particles of Tymochtee Dolomite

- Half-calcined Tymochtee dolomite at atmospheric pressure
- Half-calcined Tymochtee dolomite at 10 atmospheres pressure.

In Figure 12, the sulfidation behavior of coarse particles of half-calcined Tymochtee dolomite is shown. While the 16/18 mesh stone may be regarded as having a particle diameter of 1000 μ, the 5/6 mesh particles were too irregular to be likened to spheres and consisted of slab-like pieces of dimensions approximately 5/16 in x 1/8 in x 1/16 in. Only one piece was used as the total sample in each experiment. Comparison of the results, however, shows that sulfidation of 30% of the calcium content of the 5/6 mesh particles (4,000 μd) takes four times as long as sulfidation of the 16/18 mesh (1,000 μd) particles.

4.2.8 Stone Substrates

The sulfidation behavior of several stones selected by NEES was briefly studied at atmospheric pressure. The stones were calcined by bringing them to 1600°F at 20 $K.min^{-1}$ and passing 400 $ml.min^{-1}$ of a fuel gas mixture over the stone, containing 0.5% hydrogen sulfide.

The stones used in this test are listed in Table 16.

No differences in rate were observed in the early part of the reaction, corresponding to 30% sulfidation of a pure limestone. Differences in the later part of the reaction reflect exhaustion of unreacted calcium oxide in the calcium deficient stones. The course of the reaction is shown in Figure 13.

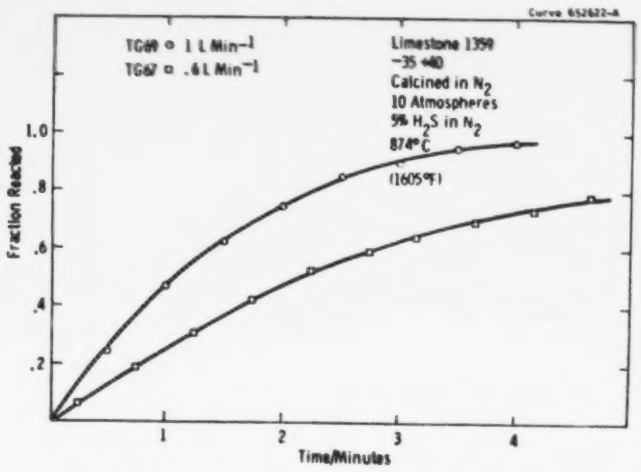

Figure 10–The Effect of Gas Flow on Sulfidation at Pressure

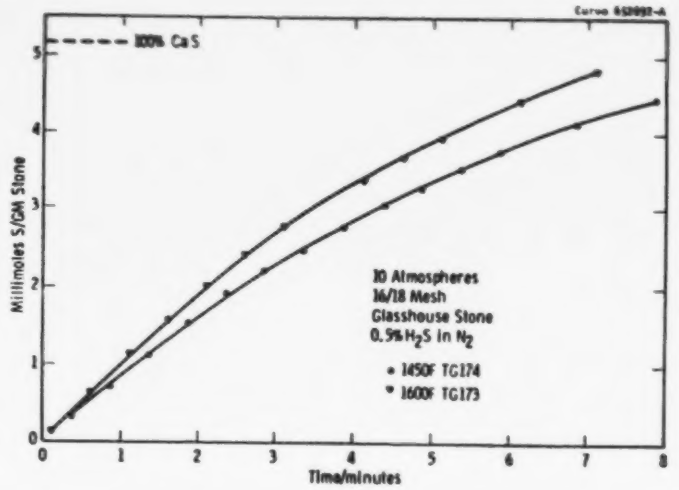

Figure 11–Effect of Temperature on Sulfidation of Calcined Dolomite

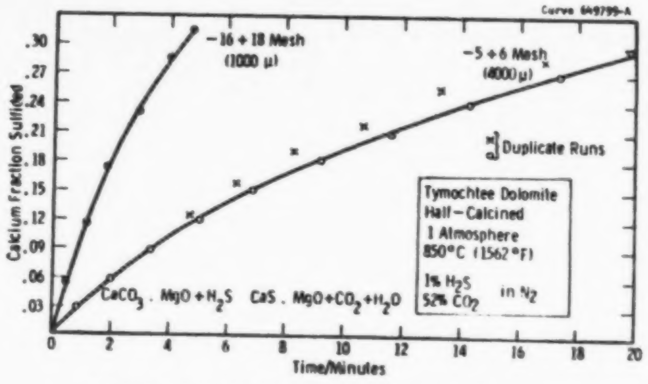

Figure 12–Sulfidation of Half-Calcined Dolomite-Atmospheric Pressure

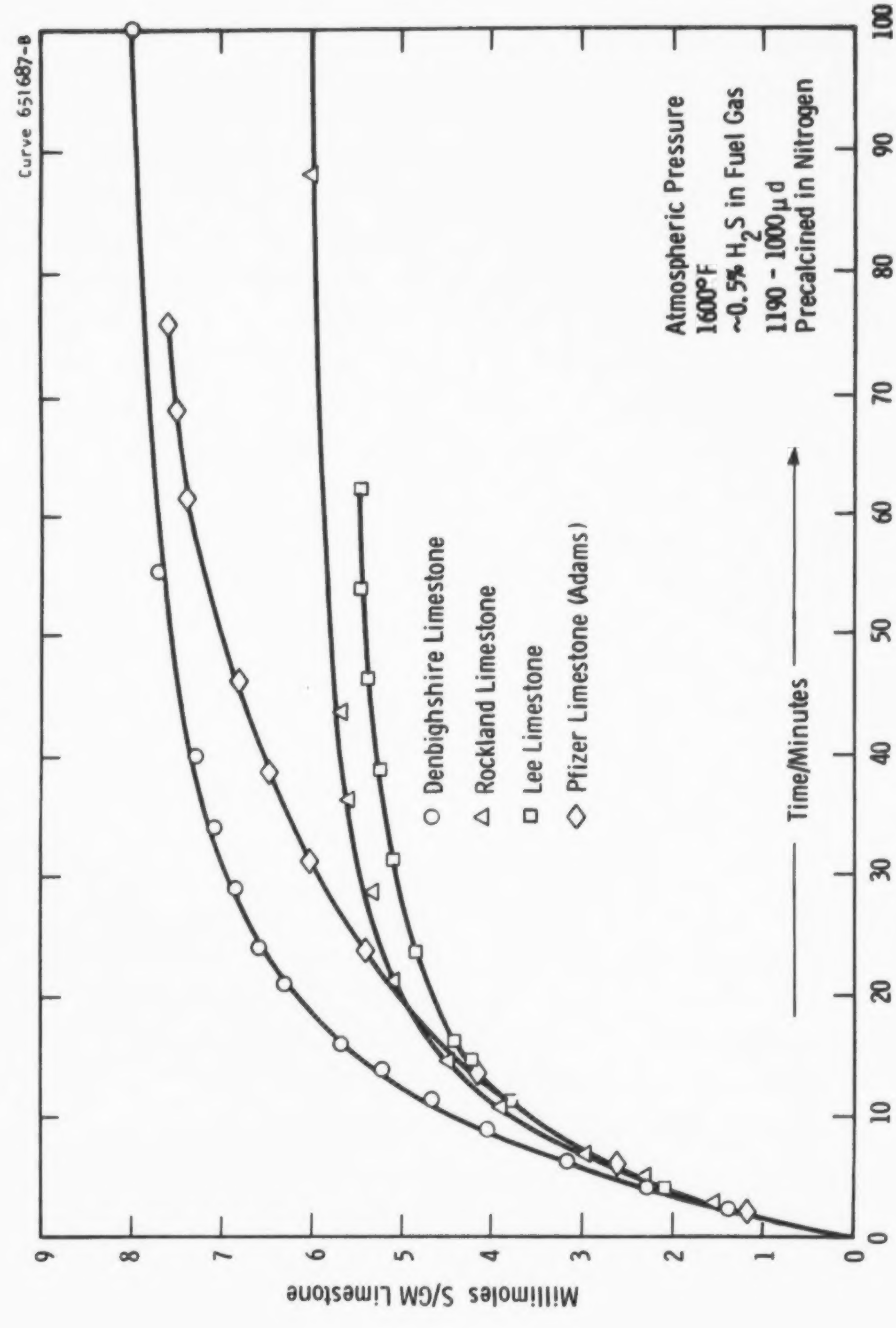

Figure 13—Sulfidation of Calcined Limestone

4.2.9 Sulfidation: Interpretation

The initial rate of sulfidation of 1.00 mm diameter 'spheres' of limestone and dolomite is virtually independent of the stone source and the purity. The effect on the rate of raising the ambient pressure to 10 atmospheres is negligible. The effect of temperature on the initial rate is slight. The calculation of mass transfer to the solids at reaction conditions is reasonably close to the experimental value. It is reasonable to conclude that mass transfer controls the rate of reaction on the TG apparatus; the rates we observe represent a minimum value which may be exceeded in other physical situations -- as, for example, fluidized beds.

The decrease in rate of reaction as the stone is used could be due to either, or a combination of, diffusion of gas through the product layer or reduction in effective calcium oxide concentration by the reduction in surface area. While distinction between the phenomena (on kinetic grounds alone) is not possible, it is possible to curve-fit the data to a conduction plus convection model and derive values for the effective diffusivity of H_2S in the product.

The sulfidation of fully calcined dolomite was studied by Pell[1]. For spheres of 6.5 mm diameter, Pell found diffusion in a product layer to participate in the rate-limiting step in sulfidation at temperatures above 750°C (1382°F). For fine powders, he concluded that "the reaction occurs homogeneously throughout the solid, provided the particle size is not too large, and diffusion in the solid will have minor importance under most anticipated conditions for applying the reaction in processes for desulfurizing fuels."

The results Ⓦ obtained at 1 atmosphere pressure are probably initially limited by gas-phase convective transfer to the solid. As reaction proceeds, the effect of an additional resistance to reaction can be seen. Pell has described this resistance as due to an effective diffusivity of H_2S (whose average value at 1742°F he found to be .056 x open space diffusivity) by curve-fitting his data to a diffusion

TABLE 16

TG EXPERIMENTS ON THE SULFIDATION OF LIMESTONES FOR NEES

Stone	Calcium Carbonate Content	Fraction Sulfided	TG No.
Denbighshire	98	.83	164
Pfizer (Adams)	96	.757	167
Pfizer (Adams)	96	.79	168
Rockland	74	.813	165
Lee	70	.776	169
Glasshouse Dolomite	52	.956	169
Aragonite (Ocean Industries)	>94	.73	175
Aragonite (Ocean Industries)	94	.23	176
Aragonite (Ocean Industries)	94	.22	177

controlled model. It is interesting to note that Hills[6] found the effective conductivity for CO_2 in lime to be .054 x open space diffusivity.

Hills concluded that CO_2 diffuses in lime by Knudsen diffusion, and therefore the diffusion is independent of the presence of other gas species. If this were true for H_2S, then pressure increases to 10 atmospheres would increase the diffusion of H_2S for as long as the Knudsen diffusion held. At the present time, the data are insufficiently precise to detect realistically such a small change.

4.2.10 Calcium Utilization

The most direct application of the Ⓦ results to fluidized bed conditions concerns the degree of utilization of calcium which can be achieved, and suggest that high utilization (80%) is possible for particles of limestone or dolomite at least as large as 16/18 mesh. This finding is in accord with the experiments reported by Consol[7] on sulfidation in fluidized beds. Consol has demonstrated the feasibility of operating even with a high-sulfur loading (80%) in the sorbent stone. Comparison of Ⓦ results with the data produced by Esso (England) in their oil gasification experiments is difficult. They obtain very good sulfur retention in the sorbent stone (> 90%), and use of models of the fluidized bed such as Koppel's predicts that this will happen. However, using Ⓦ data with Koppel's model shows that sulfur losses should not become significant until virtually all the stone has been sulfided, and the test results do not cover a sufficiently long time interval to check this prediction.

4.2.11 Fuel Gas

The fuel gas mixture used in some sulfidation experiments contained in the following gases:

	Air Products	Ⓦ
CO_2	16.4	16.9 ± 0.2
CO	10.0	10.2 ± 0.2
CH_4	2.5	2.54 ± 0.1
H_2	25	26 ± 2
N_2	balance	

These are minor differences between the suppliers' analysis and that performed at Ⓦ by the Analytical Chemistry Laboratory. No change in composition was noted over a four-month period.

Stone Sorbents	Supplier
Limestone 1359	M. J. Grove Lime Co., Stephens City, Va.
Tymochtee dolomite	courtesy of Consolidation Coal Co.
Dolomite 1337	sample courtesy of Argonne National Laboratories: Chas. Pfizer & Co., Gibsonburg, Ohio
Denbighshire Limestone	sample courtesy of S. Rakes [E.P.A.] and G. Moss [Esso (England)]
Pfizer Limestone	Pfizer, Inc., Adams, Mass.
Aragonite	Ocean Industries, Inc., Fort Lauderdale, Florida
Lee	Lee Lime Corp., Lee, Mass.
Rockland	Rockport Lime Co.
Tymochtee	C. E. Duff & Sons, Huntsville, Ohio

4.2.12 Mass Transfer Limits

Calculation of the mass transfer coefficient from H_2S in a gas stream to 1.00 mm diameter spheres, the effect of pressure on the limiting rate of sulfidation can be found.

$$\mu = 4.260 \times 10^{-2} \text{ centipoise} \qquad 0.5\% \ H_2S \text{ in } N_2$$
$$D_{12} = 1.73 \text{ cm}^2 \text{ sec}^{-1}$$

calculated according to the procedures in reference 9.

$$\frac{\alpha_{11} d}{D} = 2 + 0.69\ (Re)^{1/2}\ (Sc)^{1/3}$$

$$= 2.279 \text{ at 1 atm } 400 \text{ ml/min}^{-1}$$

$$= 2.82 \text{ at 10 atm } 2000 \text{ ml/min}$$

$$\frac{R_{10}}{R_1} = 1.24$$

(tube diameter = 1 inch
temperature = 1600°F).

4.2.13 Comparison of TG Data and Fluidized Bed Results

To predict the concentration of H_2S in the exit gas from a fluidized bed of limestone, the following calculation was made:

Fluidized Bed (Esso England Data)		Data for TG 164
2.3 wt % S in oil		0.5 % H_2S
fuel rate	247 gm/min	1600°F
air rate	617 ℓ min^{-1}	1000 μ Denbigshire
velocity	5.6 ft sec^{-1} (v)	
bed depth	15.5 13.0 inches (x)	
temp.	1561°F	
particle size = 496 average diameter.		

From this data, we derive

- $[H_2S] = .56\%$
- Bed volume ≈ 36 litres
- 25 kg limestone
- 330 gm moles CaO .

From TG data, which is approximated to a first order reaction [TG 164], the reaction rate can be written as

$$\frac{-dnH_2S}{dt} = k \ . \ CH_2S \ . \ A \ (1-\alpha)^n$$

$k = 6.07 \text{ sec}^{-1} \text{ cm}$
$A = 60 \text{ cm}^2/\text{cm}^3$
$\alpha = Ca^{2+}$ fraction sulfided.

Bed model used:

$$C = Co \ e^{-k \ A \ (1-\alpha)^{2/3} \ \tau} \qquad \tau = \frac{X}{V} = \text{gas residence time.}$$

C can be determined by calculating the value of α at finite time intervals and using the adjusted value of α in the succeeding time interval.

Using this model, the sulfur retention in the bed will not fall to 90% until 1110 minutes have passed, i.e., until virtually all the calcium oxide has sulfided. In the test run, which lasted 135 minutes, 96.5% sulfur retention in the bed was noted.

4.2.14 Outline of the Apparatus

A comprehensive description of the thermogravimetric system and its use is given in Volume I, Section 3.2 of this report. An outline of the apparatus design and capabilities is given here.

A DuPont 951 thermal balance assembly was mounted in a pressure shell similar to that described by Brown.[10] The corrosion prevention system was similar to that described by Ruth,[5] but the effluent gas was ducted away from the quartz reactor, out of the pressure shell. A cooling system prevented heating of the balance mehanism and pressure shell.

The system had the following capabilities:

- Operation at temperatures: 77°F → 2112°F
- Operation at pressure up to 30 atmospheres (15 atmospheres was the maximum pressure used in this work)
- Up to 4 litres per minute flow past the sample (in a 1 in. diameter tube) of mixtures of the gases H_2S, SO_2, CO, CO_2, N_2, H_2O, at 10 atmospheres
- Sample capacity of 1 gram (10 to 20 mg was found to be kinetically optimal for fast reactions at high temperatures)
- Noise level in the weight change record was < .04 in. at a sensitivity of .5 mg/in, when operating a $2\ell/\text{min}^{-1}$ gas flow at 10 atmospheres pressure and 1600°F.

4.3 WASTE DISPOSAL

The goal of pollution abatement has not been reached when a sulfur-rich fuel has been combusted and clean (sulfur dioxide fuel) stack gas released to the environment: the fuel sulfur remains to be disposed of in an environmentally acceptable form.

The sulfur-containing product of the fluidized-bed oil gasification/desulfurization process is a dry solid of composition CaS · CaO. The proportion of each component will depend on the operating parameters of the system (once-through, perhaps 30 to 50% CaS, or regenerative; perhaps 1 to 2% CaS make-up rate; etc.). Disposal of this solid may be accomplished by a variety of methods; the preferred method will undoubtedly reflect local circumstances and temporal economics.

There are two fundamental paths which can be chosen. The sulfur can be extracted from the CaS · CaO, or the solid can be rendered inert for dumping. Potential markets for the waste solids may develop in certain localities.

Sulfided limestone/dolomite is not inert, since its calcium sulfide component liberates hydrogen sulfide gas slowly on exposure to moist air, creating an odor problem and ultimately yielding sulfur dioxide, sulfite, and sulfate pollutants. Several processing alternatives are currently under study. Two dry processes which can be used to convert the waste into a stable solid are oxidation and sintering. The simplest and most attractive proposed scheme is air oxidation of the CaS to $CaSO_4$ (calcium sulfate). The process is dry, and the end-product is a form of calcium/sulfur which is stable to the elements over geological time periods and has some promise as an aggregate or filler material.

The calcium oxide component of the by-product may pose environmental problems, depending on its activity in hydration and carbonate formation on exposure to the atmosphere. It is not known if the residual oxide can be sealed in by the sulfide oxidation or by formation of a mist of carbonate on the surfaces (both external and internal). Further work on this aspect of the solids disposal is currently under way at Westinghouse.

Alternatively, the waste CaS · CaO could be sintered at high temperatures (> 2552°F) in air, releasing the sulfur as sulfur dioxide, and yielding dead-burned lime as the disposable solid. Sulfur dioxide would be obtained at concentrations suitable for use in the case of high-sulfur stone loading.

Where the dry process is unsuitable, wet processes may be utilized. One of the wet processes being developed is a wet slurry process which can be used to recover hydrogen sulfide, the waste calcium being converted to calcium carbonate, a solid which is environmentally stable. This process is being investigated by Consolidation Coal Company and by Westinghouse.

In this work, the dry oxidation process in which calcium sulfide is oxidized to calcium sulfate was studied. The object of the study was to establish the technical feasibility of the conversion reaction, by TG studies of the rates of reaction, extent of conversion, and boundary conditions, where such factors as loss of SO_2 prevent attainment of the goal of pollution abatement. Further study of this process is continuing at Westinghouse.

The oxidation of calcium sulfide to calcium sulfate is of interest as a means of rendering waste calcium sulfide inert for disposal purposes. It is also one of the reactions in the oxidizing/reducing cycle proposed to explain results from fluidized bed combustion experiments.[11]

The reaction

$$CaS + 2O_2 \rightarrow CaSO_4$$

is exceedingly exothermic, $\Delta H_{298} = -213$ kcal. mole.$^{-1}$ However, at high temperatures, ~2000°F, it becomes thermodynamically unfavorable with respect to the parallel oxidation reaction[12]

$$CaS + 3/2\ O_2 \rightarrow CaO + SO_2 \quad .$$

This latter reaction is undesirable in waste disposal in that it results in a net loss of sulfur at a concentration too low to be used in further processing.

The object of this investigation was to determine under what conditions calcium sulfide wastes from desulfurization processes in fuel gasification could be oxidized to inert calcium sulfates, at atmospheric pressure.

4.3.1 Conclusions from this Work

- Calcium sulfides prepared from dolomite sulfidation may be largely oxidized to calcium sulfate. The degree of conversion depends on the extent to which the original stone has been sulfided and the temperature of the oxidation reaction.
- Calcium sulfides resulting from limestone sulfidation cannot be directly oxidized in air to produce an innocuous calcium sulfate for waste disposal.

- While conversion to sulfate is more complete at higher temperatures (1472°F), there is a danger that some of the sulfur is lost as sulfur dioxide in the process. Therefore, further investigations should measure the sulfur dioxide evolution during the oxidation. This should enable a reaction temperature to be chosen which renders inert as sulfate the largest possible fraction of the fuel sulfur.
- Residual calcium sulfide in waste calcium sulfate may be eventually released to the air to form sulfur dioxide. The weathering of this material should therefore be studied in order to assess this contribution to the overall sulfur emissions.

4.3.2 Previous Investigations

A limited amount of information of use has been published. Heating calcium sulfide in pure dry air between 1472°F and 1742°F, yields a mixture of CaS, $CaSO_4$, and CaO, and sulfur dioxide is evolved. At the same time, the reaction between calcium sulfide and calcium sulfate proceeds:[13]

$$CaS + 3CaSO_4 = 4CaO + 4SO_2 \quad .$$

A Russian investigator[14] reported that air oxidation of calcium sulfide from waste (containing 5 to 20% CaS) began at 932°F. The process takes five minutes to complete with a 0.25 to 0.50 mm fraction. Heat liberated in the process accelerates the reaction until a film of oxidation product inhibits the reaction. The higher the temperature of oxidation the longer it takes the film to form. (The results refer to 2.0 gm samples heated in a tubular furnace.)

4.3.3 Experiments

Our study of this reaction falls in three parts:

- The oxidation of sulfided limestone (BCR 1359)
- The oxidation of sulfided dolomite (Tymochtee)
- The oxidation of sulfides prepared from a series of limestones.

A list of the experiments, materials, and conditions is given in Table 17.

The DuPont TG apparatus (950 or 951) was used in all experiments. Compressed air was obtained from the piped supply to the laboratory. Sulfate was determined by calcination of the stone at 2010°F on the TG apparatus. Oxidation runs were carried out, both subsequent to direct sulfidation on the TG apparatus, and on samples prepared externally, cooled, and reheated to temperature.

In the first set of experiments, air was passed over the samples at 200 ml./min^{-1} in a 1-in. diameter tube. In the later experiments the air flow was 450 ml. min^{-1} through a 2-cm diameter tube.

A complete description of TG procedures is given in Volume I, Section 3.2.

4.3.3.1 Substrates

Four different substrates were used:

- CaS from Fisher
- CaS from limestone 1359
- CaS from H_2S + CaO on the TG
- CaS from H_2S + $CaCO_3$ on the TG.

Additional runs were performed on calcium sulfite and calcium sulfate.

4.3.4 Initial Investigation

Samples of sulfided limestone (BCR 1359) were prepared by passing 15% H_2S in nitrogen over calcined limestone in a horizontal tubular furnace.

The oxidation of the product was studied on the TG apparatus.

Two kinds of experiments were performed:

- Polythermal runs: heating the sample from 700° to 2010°F
- Isothermal runs: the sample was brought to temperature under nitrogen, and air was then introduced; or the sample was abruptly brought to temperature by plunging the balance beam into the furnace at temperature.

4.3.4.1 Polythermal Runs

The course of typical polythermal runs on sulfided limestone is shown in Figure 14. Oxidation becomes perceptible below 750°F, slows down after 932°F, and increases in rate again above 1112°F, terminating at 1472°F, when the rate becomes very slow. At about 1980°F, a weight loss becomes perceptible. Continued heating results in a weight loss corresponding to stoichiometric decomposition of calcium sulfate to calcium oxide, and serves to confirm that the weight gain observed is due to formation of calcium sulfate. The total conversion noted corresponded to oxidation of 20% of the sulfide present in the stone.

Powdered calcium sulfide (Fisher) was slower to oxidize, but at temperatures above 1800°F it yielded 70% conversion to calcium sulfate.

4.3.4.2 Isothermal Oxidation

Isothermal runs were carried out at 932°F, 1292°F, and 1654°F, with sulfided limestone. Yields of sulfate observed were 3.5% in 750 minutes (932°F), 8% in 35 minutes (1292°F), and < 1% (1654°F). The course of reaction at 1292°F (TG 21) is shown in Figure 15. The sample from this experiment was cooled, lightly crushed, and reheated to 1292°F.

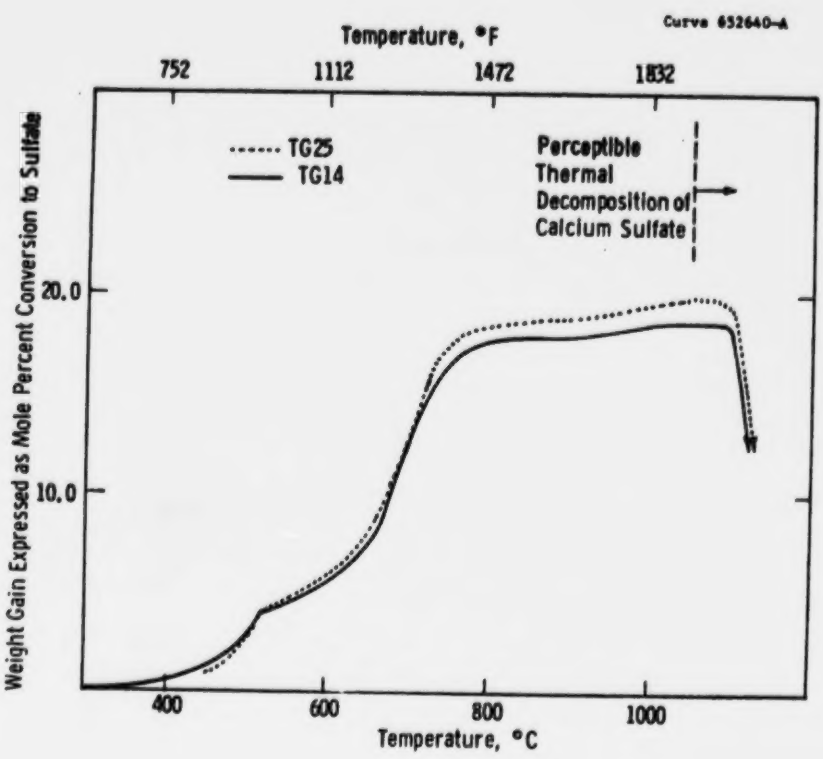

Figure 14—The Oxidation in Air of Sulfided Limestone 1359

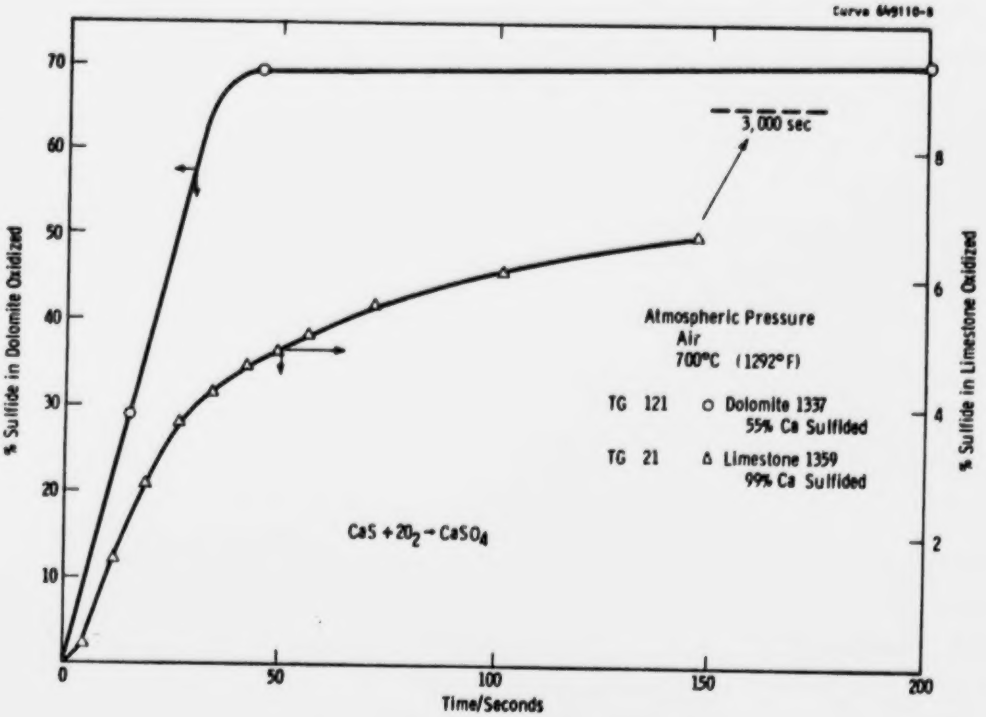

Figure 15—Oxidation of Sulfided Limestone and Dolomite

An additional 9.6% oxidation occurred before reaction essentially ceased. The temperature was then raised to 1734°F and further reaction occurred, yielding ultimately 33% conversion.

Sudden exposure of the sulfided limestone to air at 1654°F caused the yellowish stone to turn white; no detectable weight increase was noted.

These results may be interpreted as a surface phenomenon. When calcium sulfide is exposed to air, calcium sulfate is formed on the surface. Further reaction requires diffusion of oxygen across a product layer. This process is slow and is accelerated only slightly by increasing the temperature. Crushing the stone exposes fresh surfaces of calcium sulfide, thereby permitting further reaction. If this mechanistic picture is correct, then it should be easier to oxidize sulfided dolomite, since the individual calcium oxide grains are dispersed in a matrix of magnesium oxide. Sulfation experiments on calcined dolomite show that the porosity of the stone is sufficiently great to accommodate the bulk of the sulfate anion without hindering access to the unreacted stone.[15] If a similar condition exists in sulfided dolomite, the product sulfate should not block access to calcium sulfide.

4.3.5 The Oxidation of Sulfided Dolomite

A sample of dolomite, prepared by sulfiding calcined dolomite (BCR 1337) in the same manner as the samples described above, was oxidized in air on the TG apparatus. The course of the reaction is shown in Figures 15 and 16. Analysis of a portion of the sulfided material iodometrically showed that the original stone was 55% sulfided. The TG curve showed that 77% of the sulfide had been oxidized. While the batch of sulfided dolomite was not sufficiently homogeneous to attach certainty to the figures, the results appeared sufficiently encouraging to proceed with further experiments.

A particle which had been oxidized on the TG experiment was mounted in plastic resin, sectioned, and stained with lead acetate.

An outer shell showed no change, but the inner area gave the characteristic black stain of lead sulfide, indicating residual sulfide in the core of the dolomite.

4.3.5.1 Polythermal Oxidation of Sulfided Dolomite

A particle of Tymochtee dolomite, 5/6 mesh, was sulfided on the TG apparatus at 1600°F in 1% hydrogen sulfide/nitrogen/CO_2 mixture. When 25% of the calcium content had sulfided by reaction of H_2S with the calcium carbonate, reaction was interrupted and the sample was cooled. Air was passed over the sample as it was heated.

The rate of oxidation observed became appreciable in the range 750° to 930°F. It then became slower at 1112°F, either because the sulfide was no longer accessible to the oxygen or because loss of carbon dioxide from the unreacted stone began to interfere. Probably both mechanisms were operative.

4.3.5.2 Isothermal Oxidation

Isothermal runs were performed at temperatures from 1022°F to 1472°F on partially sulfided dolomite. The experimental conditions and the results are shown in Table 18. The course of reaction is shown in Figures 16 and 17. Initially, reaction was extremely rapid in all cases, but eventually the reaction rates fell off into a very slow mode of oxidation. The degree of oxidation attained before the reaction became slow increased with temperature; at 1022°F, reaction essentially ceased after 16% oxidation; at 1202°F, over 60% oxidation occurred; and at 1472°F, 92% conversion was noted.

Sectioning the 932°F sample and staining it with lead acetate showed that residual sulfide was concentrated in a thin band at the outer edges of the stone.

Using a smaller particle size [(16/18) mesh-(1000 μ diameter)] sample, the sulfate yield was 85% at 1382°F, and 75% at 1292°F. Increasing the temperature to 1562°F did not increase the conversion in the latter case, and a slight weight loss occurred, probably due to temperature overshoot in the furnace when the temperature was changed.

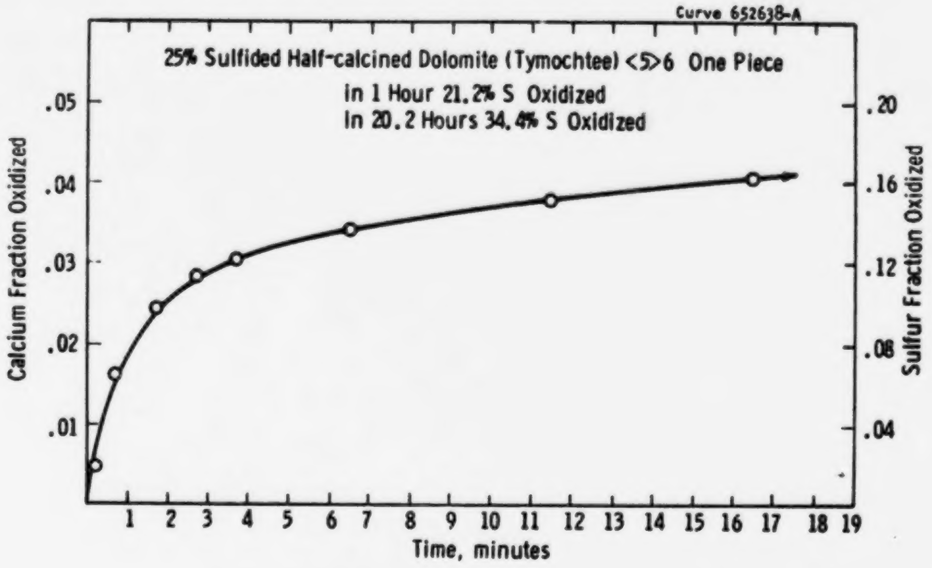

Figure 16—Oxidation of Dolomitic Calcium Sulfide at 550°C (1022°F)

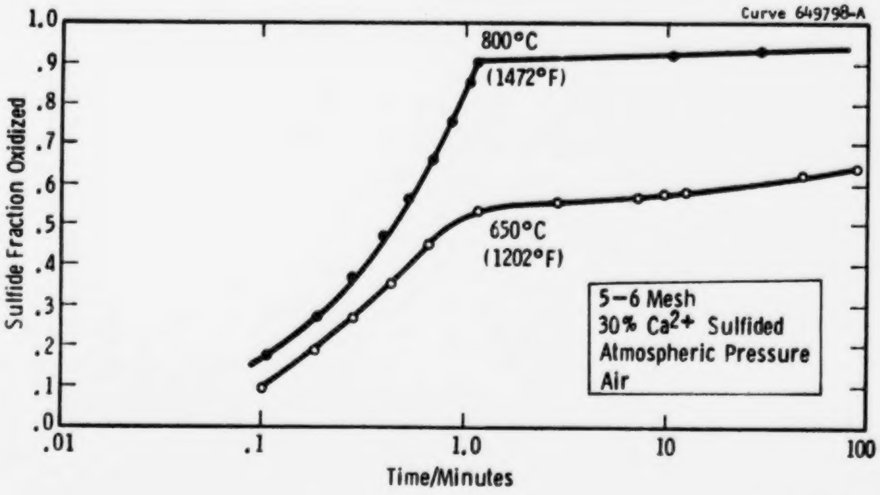

Figure 17—Oxidation of Sulfided Tymochtee Dolomite

TABLE 17

THE OXIDATION OF SULFIDED DOLOMITE

Stone	Size	Fraction Sulfided	Sulfide Fraction Oxidized	Temp. (°C)	Comment	TG No.
Tymochtee Dolomite	5/6	.166	.20	200-700 (392-1292°F)	polythermal	128
Tymochtee Dolomite	5/6	.29	.20	550 (1022°F)	residual sulfide near surface	129
Tymochtee Dolomite	5/6	.24	∼.60	650 (1202°F)	calcination interferes with weight recording	130
Tymochtee Dolomite	5/6	.305	.64	650 (1202°F)	fully calcined before oxidation	131
Tymochtee Dolomite	5/6	.297	.92	800 (1472°F)	temperature rise noted on TG record	132
Tymochtee Dolomite	16/18	.297	.87	750 (1382°F)	sulfate assessed by thermal decomposition	133
Tymochtee Dolomite	16/18	.31	.70	700 (1292°F)	temperature increase to 850°C (1562°F) does not complete oxidation	134

TABLE 18

TG EXPERIMENTS ON THE OXIDATION OF SULFIDED LIMESTONES AND DOLOMITES

Stone	Calcium Carbonate Content	Fraction Sulfided	Sulfide Fraction Oxidized	Comment	TG No.
Denbighshire	98	.83	.12		164
Pfizer (Adams)	96	.757	.12		167
Pfizer (Adams)	96	.79	.10	oxidation at 1600°F	168
Rockland	74	.813	.195		165
Lee	70	.776	.194		169
Glasshouse Dolomite	52	.956	.70		166
Aragonite (Ocean Industries)	>94	.73	.30		175
Aragonite (Ocean Industries)	94	.23	.595		176
Aragonite (Ocean Industries)	94	.22	.533	sulfidation in the absence of fuel gas	177

The 1382°F sample was heated to 2012°F and the sulfate was decomposed thermally. The weight loss for conversion to calcium oxide agreed with the assumed sulfate formation, within 1%.

4.3.6 The Oxidation of Sulfides Prepared from a Series of Limestones

A series of air oxidation runs was carried out on samples of limestone submitted by NEES for testing. The samples were calcined, sulfided in a fuel gas at atmospheric pressure and 1600°F until sulfidation became very slow, and then oxidized, after cooling to 1500°F (816°C). The results are summarized in Table 19. With one exception, oxidizability of the product sulfide increased with decreasing calcium content of the original stone. With pure limestone, 12% conversion of sulfide to sulfate was possible. With a dolomite substrate, 70% conversion was noted. Increasing the reaction temperature to 1600°F had no marked effect on the yield.

One sample, an aragonite (Ocean Industries) proved anomalous in that it contained 96% $CaCO_3$, but 24% of its sulfur content after sulfidation was oxidizable. Further experiments were conducted in which the stone was sulfided to less than one-third of its capacity. In one run, hydrogen sulfide in fuel gas sulfided the stone. In a second run, hydrogen sulfide in nitrogen sulfided the stone. The sulfate yields on oxidation were 59% and 54%.

4.3.7 The Kinetics of Oxidation

The reaction between a gas and solid which produces a product shell often leads to diffusional limitation of the reaction rate. In the case of pure limestones, the diffusion process is extremely slow, permitting about 1% hr^{-1} oxidation at the 7% oxidized level at 1292°F However, the initial rate of reaction is extremely rapid ($\sim$ 1% oxidation in 7 seconds). Since the reaction is extremely exothermic, ($\sim$ 220 kcal $mole^{-1}$) our experimental measurements are complicated by the non-isothermal nature of reaction. Calculation of the self-heating, based

TABLE 19

CALCIUM SULFIDE OXIDATION: TG EXPERIMENTS

TG No.	Sample Source	Mesh	Temp. Range °F	Atmosphere	Mole Percent Sulfate Yield
14	1359		70-2140	air	Fig. 14
15	Fisher		70-2140	air	
17	Fisher		1560	O_2	8.8%
18	Fisher		70-2140	O_2	61.0%
19	Fisher		70-2140	O_2 (dry)	74%
20	1359		932	air	4%
21	1359		1292	air	8.6%
22	Fisher		70-2140	O_2 (H_2O)	30%
23	1359		70-1292		23%
24	$CaSO_4$ 2·H_2O		70-2140		
25	1359		70-	air	18.5%
26	TG 21		70-1400-1733	air	
27	1359		Shock 1742	air	2%
28	1359		Shock 1760	air	2%
29	1359		70-2102	air less CO_2	9%
30	$CaSO_3$		70-2140	air	
31	$CaSO_3$		70-2140	air	Appendix
32	$CaSO_3$		70-2140	air	
33	$CaSO_3$		70-1868	air	
34	1359		1300	air	9%
35	1359		Shock 1292	air	4%
40	1359		70-1652	air	15%
43	1359		70-1148	air	15%
121	1337			air	77%
128	Tymochtee	-5/+6	1300	air	
129	Tymochtee	-5/+6	392-1292	air	Table 17
130	Tymochtee	-5/+6	1022	air	
131	Tymochtee	-5/+6	1202	air	
132	Tymochtee	-5/+6	1202	air	
133	Tymochtee	-16/+18	1472	air	
134	Tymochtee	-16/+18	1382	air	
161	Glasshouse	16/18	1292; 1562	air	
162	Glasshouse	16/18	1500	air/N_2	Table 18
164	Denbighshire	16/18	1500	air/N_2	
165	Rockland	16/18	1500	air	
166	1337	16/18	1500	air	
167	Pfizer	16/18	1500	air	
168	Pfizer	16/18	1500	air	
169	Lee	16/18	1600	air	
175	Aragonite (Ocean Industries)	16/18	1500	air	
176		16/18	1500	air	
177		16/18	1500	air	

on a simple Semenov model show that 500μd particles may be subject to excess temperatures of 212°F or more in the TG experiment. In the oxidation of sulfided dolomite (TG 122), a temperature increase was noted when air was admitted to the sample. Since the thermocouple was in a quartz shield and was adjacent to but not in contact with the sample, the actual temperature of the sample must have been much greater than that indicated by the recorded temperature rise of 63°F (Figure 18). Since the weight versus time curves recorded are not the result of isothermal reaction, it would be unwise to extract kinetic parameters from the data. Before this can be done, temperature measurements of the solid surface during reaction are required.

The polythermal experiment on the oxidation of dolomitic calcium sulfide shows an acceleration of rate with temperature in the range of 392°to 1022°F Relatively little material is oxidized in this range, so the influence of sample geometry of reactant depletion should be small. The apparent activation energy for reaction in the range 0 to 3% is 28 kcal mole^{-1}. There is no independent evidence for this value.

4.3.8 Summary

Disposal of sulfide-containing dolomites by air-oxidation of the calcium sulfide to calcium sulfate will be possible only with a very low-sulfur loading in the sorbent. Grinding the material before oxidation will increase the sulfur-load for which the oxidation treatment can be successful. The process will be an important option for disposal of spent sorbent from regenerative systems.

Problem areas in which further experimental work is required include investigation of the availability of the residual lime in the stone for reaction with the environment, and determination of the activity of sulfide in incompletely oxidized sorbent. Fluidized bed studies of the reaction are recommended to confirm the process parameters outlined by the TG experiments. Alternative spent stone processing systems -- both wet and dry -- should also be developed. Further study on alternative

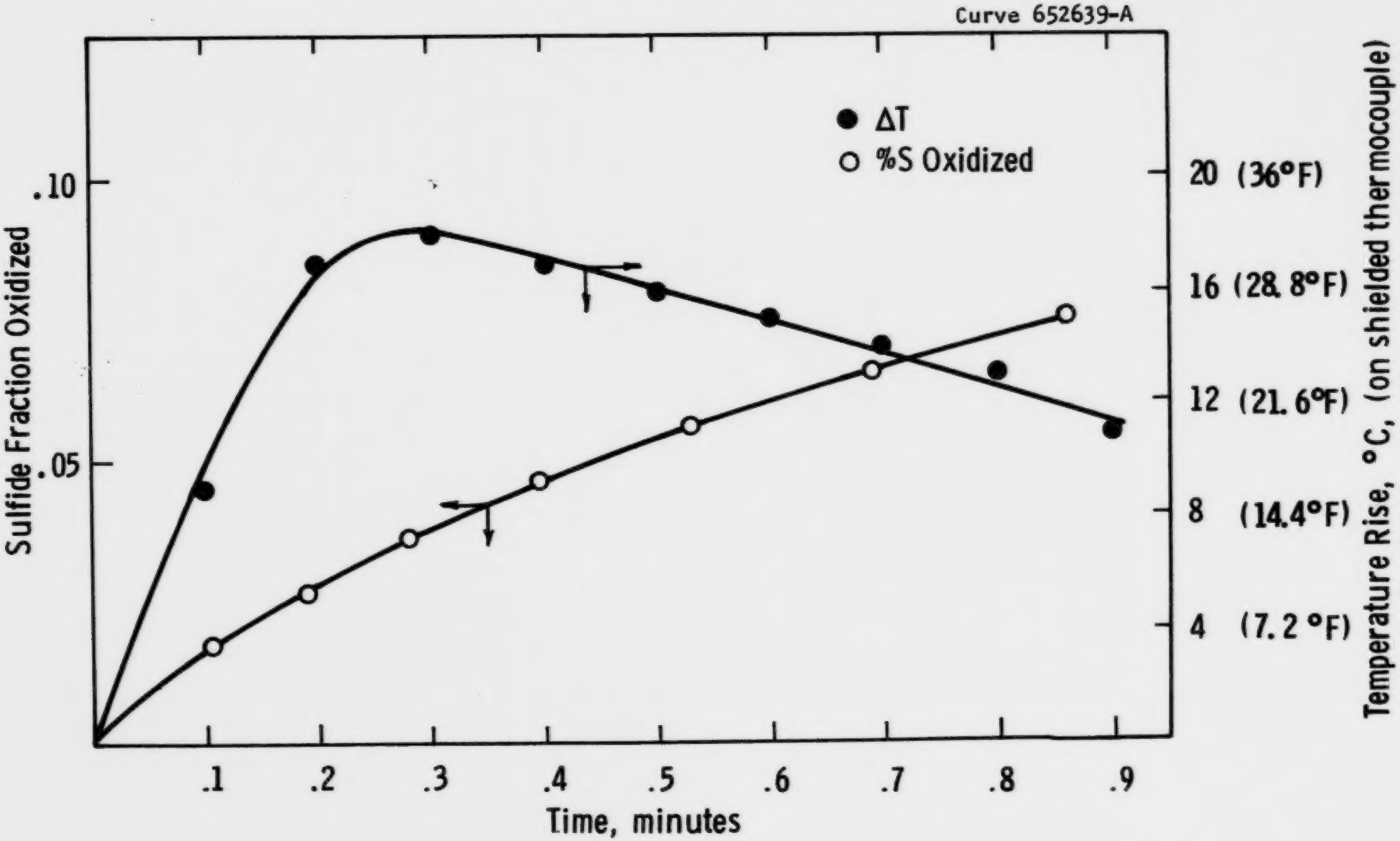

Figure 18—Oxidation of Dolomitic Calcium Sulfide at 800° C (1472°F)

disposal systems are currently under way at Westinghouse as a support effort to the demonstration plant program.

4.3.9 Some Implications of the TG Study on Calcium Sulfide Oxidation

The disposal of waste calcium sulfide raises the question of an additional contribution to the overall sulfur plant emissions. For conditions where insufficient oxidation of sulfide to sulfate is attained, the build-up of the rate of sulfur emission will depend on the rate at which sulfur is released from a growing waste pile containing some calcium sulfide.

Sulfur, released by weathering of the waste, would probably consist initially of hydrogen sulfide which would oxidize in the atmosphere to sulfur dioxide. The implications of the waste emissions, as sulfur dioxide, is considered here.

It is recognized that other criteria will apply to the waste material, and that environmental criteria relating to odor and solubility of the waste may place other restrictions on the system.

If the total emissions from the waste and the plant are together to conform to the federal standard of 0.8 lb $SO_2/10^6$ Btu, then the fraction of waste calcium sulfide which must be rendered inert, for a given fraction of sulfur retention in the fluidized bed, can be calculated. Figure 19 shows how total emission may be held below the permissible limits. For example, if 90% of the fuel sulfur is captured in the fluidized bed, then oxidation of 85% of the sulfide in waste sulfided limestone or dolomite ensures that total emission meets the legal requirement (if all sulfur losses arise by weathering of calcium sulfide). It should be noted that 92% oxidation of sulfide waste has been achieved in TG experiments.

Because of waste accumulation, insufficient oxidation of calcium sulfide would lead to an increasing sulfur evolution from the waste pile, which would gradually rise to exceed the emission standard, assuming a first order leaching process. For example, if 90% of the fuel sulfur is

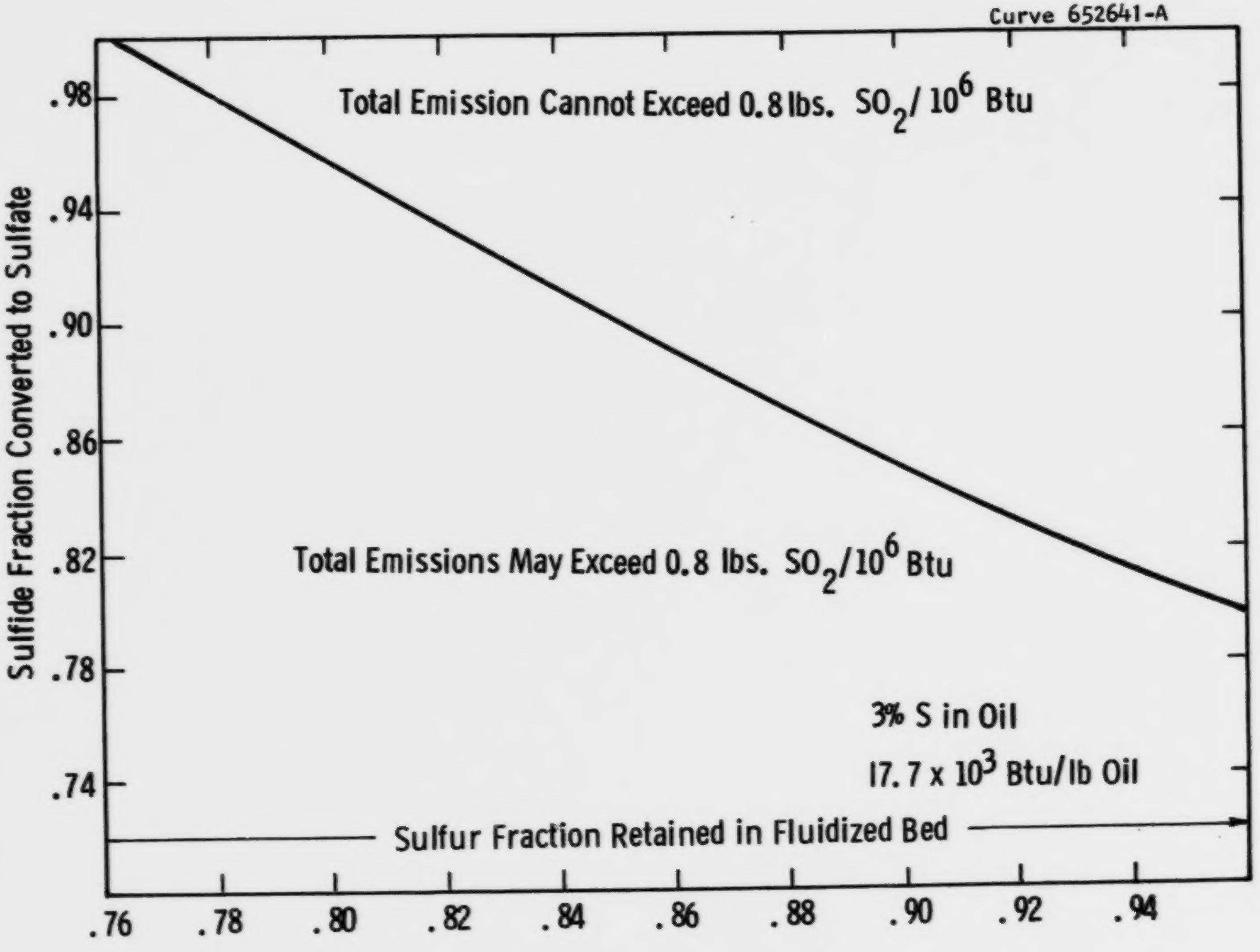

Figure 19—Waste Oxidation Required for EPA Standards

retained in the fluidized bed and 70% conversion of waste sulfide to sulfate is attained, then the total emissions would exceed the federal standard at a 6% sulfur annual loss rate from the waste after 12 years.

It is not possible to determine loss rates of this order (.06 yr^{-1}) in a simple laboratory experiment, since it corresponds to a yield of 1 μg per day of H_2S from a 50 gram sample of oxidized, sulfided dolomite.

5. REFERENCES

2. ATMOSPHERIC-PRESSURE FLUID BED OIL GASIFICATION

1. Evaluation of the Fluidized Bed Combustion Process, Westinghouse report to Office Air Programs, EPA, November 1971.

2. Newby, R. A., D. L. Keairns and D. H. Archer, "Assessment of Fluidized Bed Oil Gasification for Power Generation," paper presented at Air Pollution Control Association Annual Meeting, June 1972; JAPCA 23 (1), 23 (1973).

3. Craig, J. W. T., G. L. Johnes, G. Moss, and J. H. Taylor, "Study of Chemically Active Fluid Bed Gasifier for Reduction of Sulfur Oxide Emissions," Interim Report, APCO Contract CPA 70-46, Esso Research Centre, Abingdon, Berkshire, England, August 1970.

4. "Study of Chemically Active Fluid Bed Gasifier for Reduction of Sulfur Oxide Emissions," Final Report APCO Contract CPA 70-46, Esso Research Centre, Abingdon, Berkshire, England, June 1972.

5. Keairns, D. L., D. H. Archer, R. A. Newby, W. L. Wright, and E. J. Vidt, "Location of Utility Partner for Atmospheric Pressure Fluidized Bed Oil Gasification Demonstration Plant," Research Memo 73-9E3-FBCOM-M1, February 1973.

3. OIL GASIFICATION FOR COMBINED CYCLE POWER GENERATION

1. Study of the Future of Low-Sulfur Oil for Electric Utilities. Hittman Associates. February 1972 (PB200257).

2. Availability of Residual Fuel Oil. M. W. Kellogg Co. Report to EPA, December 1970, PB 198817.

3. Hauser, L. G.. "Optimum Uses of Energy Sources." Paper presented at Spring Conference, Southeastern Electric Exchange, New Orleans, La., April 1971.

4. Study of Chemically Active Fluid Bed Gasifier for Reduction of Sulfur Oxide Emissions. Esso Research Center, Abingdon, England. OAP Contract CPA 70-46, June 1972.

5. D. L. Keairns, D. H. Archer, R. A. Newby, W. L. Wright, E. J. Vidt. "Location of Utility Partner for Atmospheric Pressure Fluidized Bed Oil Gasification Demonstration Plant." Research Memo: 73-9E3-FBCOM-M1, February 19, 1973.

6. Moss, G.. "The Desulfurizing Combustion of Fuel Oil in Fluidized Beds of Lime Particles." Paper presented at First International Conference on Fluid Bed Combustion, Hueston Woods, Ohio, November 1968.

7. Craig, J. W. T., G. L. Johnes, G. Moss, and J. H. Taylor. "Study of Chemically Active Fluid Bed Gasifier for Reduction of Sulfur Oxide Emissions." Interim Report, APCO Contract CPA 70-46, Esso Research Centre, Abingdon, Berkshire, England, August 1970.

8. Newby, R. A., D. L. Keairns, and D. H. Archer. "Assessment of Fluidized Bed Oil Gasification for Power Generation." Paper presented at the 65th Annual meeting of the Air Pollution Control Assoc., Miami Beach, Florida, June 1972.

9. Information provided during discussions with Shell, May 3, 1972.

10. Information provided during discussion with Texaco, May 24, 1972.

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4. SUPPORT STUDIES

1. Pell, Melvyn. "Reaction of Hydrogen Sulfide with Fully Calcined Dolomite." Ph.D. Thesis, The City University of New York, 1971.

2. Kotthoff, I. M. and P. J. Elving. "Treatise on Analytical Chemistry," 4, Pt. 2, p. 144, Wiley, New York, 1966.

3. "Methods of Analyzing and Testing Coal and Coke." U.S. Bureau of Mines, Bulletin 638, 1967.

4. Wise, A. O.. "Proc. 2nd International Conference on Thermal Analysis," ed. Schwenker, R. F., Jr., and P. D. Garn. Academic Press, New York, 1969, p. 1095.

5. Ruth, L. A.. "Reaction of Hydrogen Sulfide with Half-Calcined Dolomite." Ph.D. Thesis, The City University of New York, 1972.

6. Campbell, F. R., A. W. D. Hills, and A. Paulin. Chem. Eng. Sci. 25, 929, 1970.

7. Curran, G.P., B. Pasek, M. Pell, E. Gorin. Monthly Progress Report No. 21, from Consolidation Coal Co. to EPA, Contract EHSD-71-15, December 12, 1972.

8. Esso Research Centre (England). Appendix D to Final Report to EPA; OAP Contract CPA 70-46.

9. Reid, R. C. and T. K. Sherwood. "The Properties of Gases and Liquids." McGraw Hill, New York, 1966.

10. Brown, H. A., Jr., Penski, E. C., and Callahan, John J.. Thermochim, Acta 3, 271, 1972.

11. Jonke, A. A., et al. "Reduction of Atmospheric Pollution by Application of Fluidized Bed Combustion." Argonne National Laboratory, Monthly Progress Report No. 23, Sept. 23, 1970.

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12. High Temperature Properties and Decomposition of Inorganic Salts. Part 1, Sulfates, NSRDS-NBS-7, 1966.

13. Gmelins Handbuch Der Anorganischen Chemie System 28, Part B, Vol. 2, p. 643.

14. Akodus, V. Ya.. Izr. Vyssh. Ucheb. Zaved., Les. 2L, (4) 96, 1969, Chem. Abs. 72, 24244.

15. Borgwardt, R. H. and R. D. Harvey. Environ. Sci. Tech. 6, 350 (1972).

APPENDIX A

PERFORMANCE OF THE OIL GASIFICATION POWER PLANT AS A FUNCTION OF THE AIR/FUEL RATIO

The effects of the gasifier air/fuel ratio on the gasification process has been estimated so that the rate of heat generation in the gasifier might be predicted to indicate the temperature control requirements in the gasification/desulfurization system. The performance of the gasification system relative to the air/fuel ratio is also considered. Approximate relationships developed serve as a basis for estimating the general behavior of the system and are not expected to be accurate descriptions of specific phenomena.

GASIFIER PRODUCT CO/CO_2 RATIO

Essentially two regions exist in the atmospheric-pressure oil gasifier. In the lower region of the bed, carbon (which has deposited on the lime) and a fraction of the incoming fuel are burned to supply heat to the gasifier. In the upper region, above the fuel injection point, H_2S absorption takes place. It is assumed that the most important reactions which determine the CO/CO_2 ratio in the gasifier product are

$C + 1/2\ O_2 \rightleftarrows CO$[*] oxidizing region of bed (A-1)

$H_2 + 1/2\ O_2 \rightleftarrows H_2O$ oxidizing region of bed (A-2)

$CO_2 + H_2 \rightleftarrows CO + H_2O$ reducing region of bed. (A-3)

If reaction A-3 is assumed to be in equilibrium in the upper, reducing region of the bed, then

---

[*] Equilibrium calculations indicate that CO is preferentially formed over CO_2. Esso (England) Progress Report No. 8.

$$\frac{M_{CO_2}\, M_{H_2O}}{M_{CO_2}\, M_{H_2}} = K_E \quad , \qquad \text{(A-4)}$$

where M_i is the molar flow rate of species i and K_E is the equilibrium constant for reaction A-3. If the total molar flow rate of oxidized carbon, CO + CO_2, is given by

$$M_{CO} + M_{CO_2} \equiv K_1 \quad , \qquad \text{(A-5)}$$

and if it is assumed that all the incoming oxygen, $M_{O_2}(I)$, is used completely to oxidize either carbon or hydrogen, then a material balance gives

$$M_{CO_2} + 1/2\, M_{CO} + 1/2\, M_{H_2O} = M_{O_2}(I) \equiv K_2 \quad . \qquad \text{(A-6)}$$

Also, if it is assumed that only CO and H_2O are formed in the oxidizing region of the bed and all the CO_2 and H_2 arise from reaction A-3 in the reducing region of the bed, then

$$M_{H_2} = M_{CO_2} \quad . \qquad \text{(A-7)}$$

From equations A-4, 5, 6 and 7 the ratio M_{CO}/M_{CO_2} may be found to yield

$$\frac{M_{CO}}{M_{CO_2}} = \frac{2\,(K_E-1)}{=2K + [4K^2 + 8\,(2K-1)\cdot(K_E-1)]^{1/2}} - 1 \quad . \qquad \text{(A-8)}$$

where $K \equiv K_2/K_1$. To estimate K it is assumed that all of the Conradson carbon in the residual fuel oil is laid down on the gasifier lime and is oxidized to form CO; and the remaining O_2 is distributed between C and H_2 in the ratio $C{:}H_2 = 2{:}1$, as in C_nH_n hydrocarbon compounds, to form CO and H_2O. Then, with a fuel having 9.3 wt % Conradson carbon, as with the U.K. fuel investigated by Esso (England), these two assumptions lead to

$$K_1 = M_{CO} + M_{CO_2} = \frac{0.093}{12} + \left[M_{O_2}(I) - \frac{0.093}{2(12)}\right] \cdot \frac{2}{3} \cdot 2 \tag{A-9}$$

$$= 0.00775 + [M_{O_2}(I) - 0.0039]\ 1.33$$

or

$$K = \frac{M_{O_2}(I)}{0.00255 + 1.33\ M_{O_2}(I)} \quad . \tag{A-10}$$

The results of equations A-8 and A-10 are shown in Figures A-1 and A-2. Figure A-2 shows the percent fuel carbon oxidized against the air/fuel ratio, and Figure A-1 shows the ratio M_{CO}/M_{CO_2} versus the air/fuel ratio. The predictions are in general agreement with the experimental findings of Esso (England) where ∿40% carbon oxidation and CO/CO_2 ∿1 has been found, though the data points are not numerous enough to merit comparison with the predictions.

HEAT GENERATION RATE

Having the relationships in equations 20 and 22, the total rate of heat production by fuel combustion may be estimated, assuming that the combustion reactions dominate over the other system reactions in terms of heat production, which is seen to be the case. Then

$$\Delta H_G = -12.0 \frac{K_1}{\left(1 + \frac{M_{CO}}{M_{CO_2}}\right)}(14{,}087) - 12.0 \frac{K_1 \frac{M_{CO}}{M_{CO_2}}}{\left(1 + \frac{M_{CO}}{M_{CO_2}}\right)}(3{,}960)$$

$$- 2.0 \left[2K_2 - \left\{\frac{2K_1}{1 + \frac{M_{CO}}{M_{CO_2}}} + \frac{K_1 \frac{M_{CO}}{M_{CO_2}}}{1 + \frac{M_{CO}}{M_{CO_2}}}\right\}\right](51{,}571),\ \frac{\text{Btu}}{\text{lb fuel}} \tag{A-11}$$

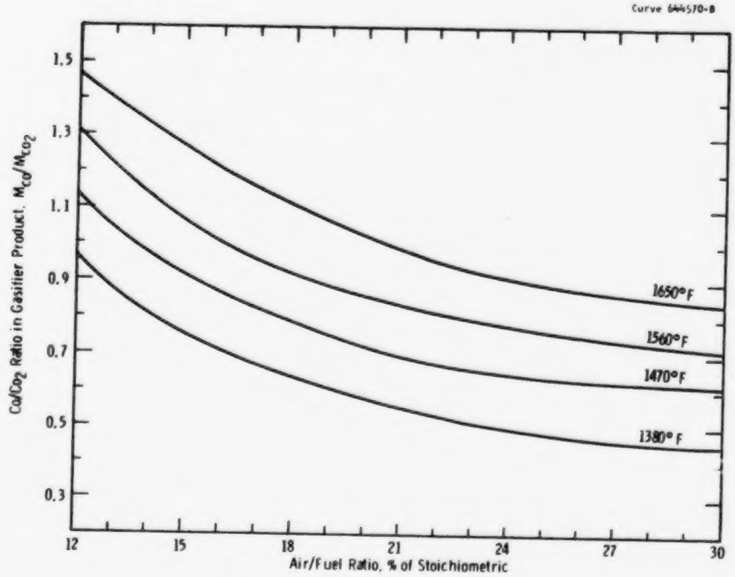


Figure A-1–Analysis of Gasifier Product Co/Co_2 Ratio

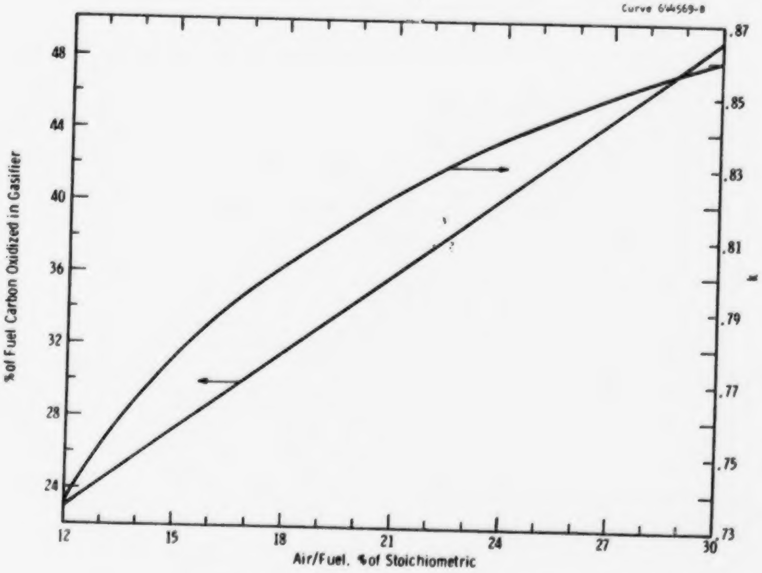


Figure A-2–Analysis of Fuel Combustion in the Gasifier where M_{CO}, M_{CO_2}, and M_{H_2O} have been expressed in terms of K_1, K_2, and M_{CO}/M_{CO_2}. This reduced to

$$\Delta H_G = -206{,}284\ K_2 + 37{,}240 \frac{K_1}{1 + \frac{M_{CO}}{M_{CO_2}}} + 55{,}622 \frac{K_1 \frac{M_{CO}}{M_{CO_2}}}{1 + \frac{M_{CO}}{M_{CO_2}}} \qquad \text{(A-12)}$$

and is compared to the data of Esso (England) at temperatures of 1562°F and 1652°F in Figure A- 3. The figure shows good agreement between the experiment and equation A-12. Having this estimate of the heat generation rate during gasification/desulfurization, the temperature control requirements for maintaining a 1600°F gasifier temperature may be found. Representing the net amount of heat generated in the gasifier by $\Delta H_{control}$, a quantity independent of the temperature control method or heat losses, then

$$\Delta H_{control} = \Delta H_G - F_p \bar{C}_p\ \Delta T_p \quad . \qquad \text{(A-13)}$$

Where F_p is the mass flow rate of the product from the gasifier per lb of fuel $\bar{C}_p$ is the mean heat capacity of the product and $\Delta T_p = 1523°F$. F_p and $\bar{C}_p$ may be expressed as a function of the air/fuel ratio to obtain the results shown in Figure A-4.

In a regenerative operation with 1% heat losses from the gasifier to the surroundings and a make-up rate of 1 mole CaO/mole sulfur, the total energy lost from the gasifier will be

$$177 + 3790\ Y - 90\ N_G \quad , \qquad \text{(A-14)}$$

and in a once-through operation with 1% heat losses to the surroundings the total heat lost will be

$$177 - 1.409 \times 10^4\ X_c N_G \quad . \qquad \text{(A-15)}$$

These losses are evaluated for $Y = 0.03$, $X_c = 0.003$, and regenerator temperature control with $N_{MR} = 0$ and $N_G = N_{G\ MIN}$ (or by heat transfer surface) to give the thermal balance air/fuel ratios shown in Figure A-4.

Curve 644566-A

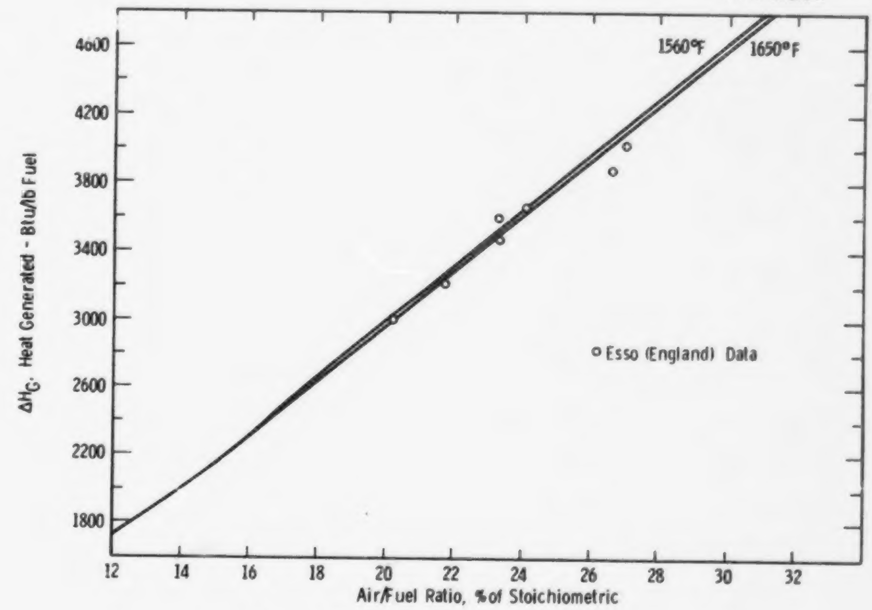


Figure A-3—Prediction of Heat Generation Rate

Curve 644572-B

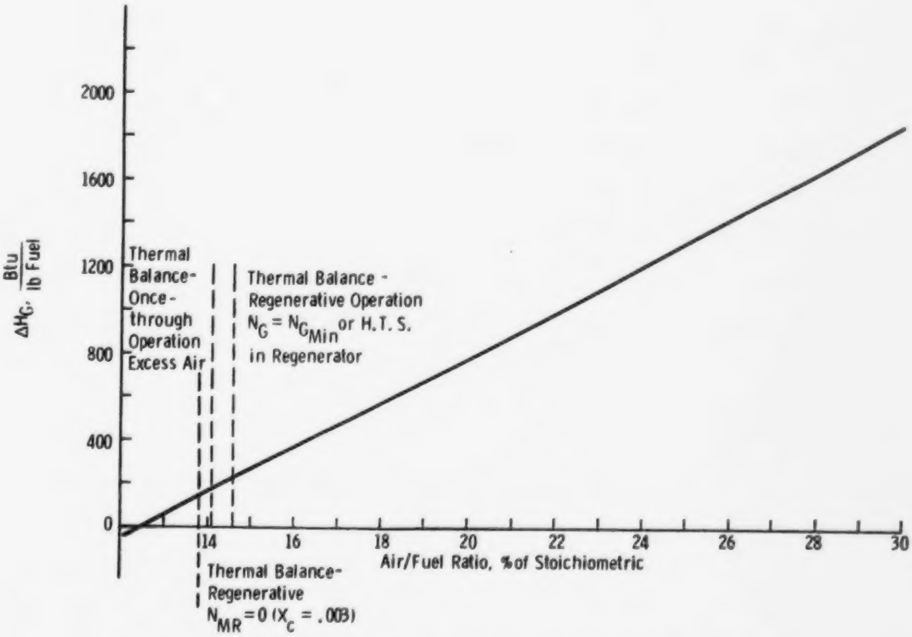


Figure A-4—Heat Generation Rate in Gasifier

These correspond to the minimum air/fuel ratios at which no additional temperature control is needed to maintain the 1600°F gasifier temperature. The thermal balance point then depends on the direct heat losses to the surroundings, the mode of operation, the temperature control method being used in the regenerator or sulfate generator, and the amount of carbon deposited on the gasifier lime, X_c.

Based on the results shown in Figure A-4, the temperature control stream required as a function of the air/fuel ratio with stack gas recycle, steam injection, or water injection is found and is shown in Figure A-5. The results are based on stack gas at 300°F with an enthalpy change of 10,400 Btu/mole in going from 300°F to 1600°F, steam at 212°F with an enthalpy change of 707 Btu/lb, and water at 77°F with an enthalpy change of 1800 Btu/lb. Results are indicated for regenerative and once-through operations with regenerator temperature control by stone circulation ($N_{MR} = 0$), make-up stone ($N_G = N_{G\ MIN}$) or heat transfer surface (H.T.S.), and sulfate generator temperature control by excess air circulation or heat transfer surface. The requirements with stack gas recycle or steam injection appear excessive at air/fuel ratios greater than 20% of stoichiometric, while great losses in thermal efficiency occur with the steam and water injection temperature control methods due to the lost heat of vaporization of water. Thus, steam and water injection should be considered only at low air/fuel ratios (< 20%).

The amount of carbon deposited on the gasifier lime and the method of regenerator temperature control have a great effect on the thermal efficiency of the gasifier. This behavior is demonstrated in Figure A-6 for regenerative and once-through operations with gasifier temperature control by stack gas recycle, heat transfer surface, or operation at the minimum air/fuel ratio (the thermal efficiency with these schemes being independent of the air/fuel ratio). Assuming 1% heat losses, 95% sulfur removal with a 3 wt % sulfur fuel, a make-up rate of 1 mole of CaO/mole sulfur fed in the regenerative case, and a limestone feed rate of 3 moles CaO per mole sulfur fed in the once-through case, the thermal efficiency, ε_T, is given by

Curve 644568-B

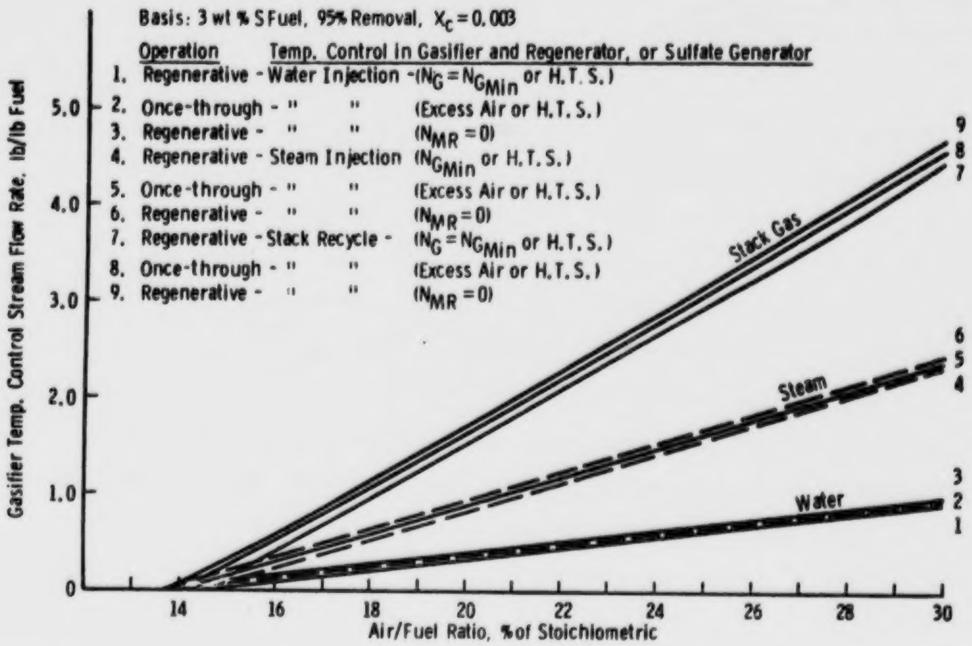


Figure A-5—Gasifier Temperature Control Stream

Curve 644564-A

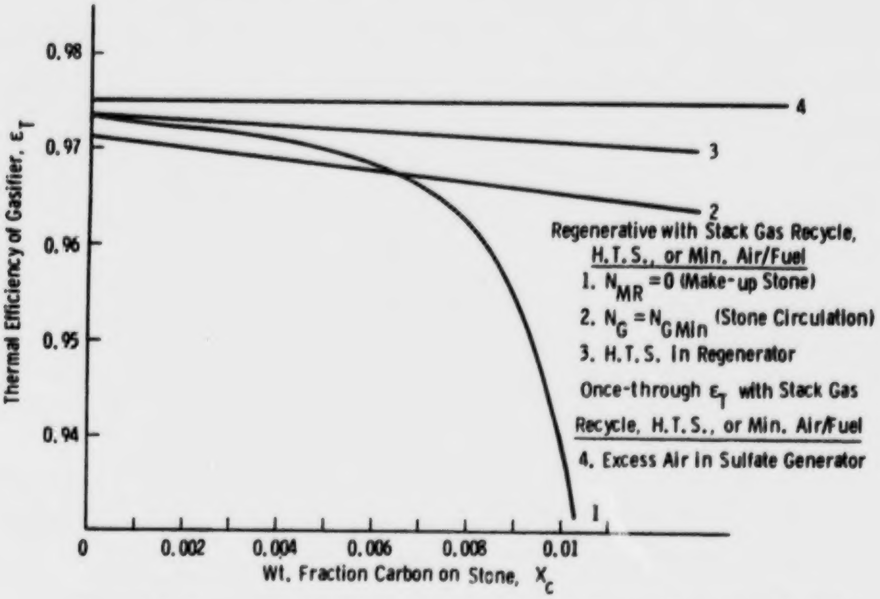


Figure A-6—Thermal Efficiency versus Carbon on Stone for Proposed Temperature Control Schemes

$$1-\varepsilon_T = \frac{\varepsilon_s Y\,(10{,}630) + X_c N_G\,(14{,}093) - 90\,N_G + 3217\,Y + 177}{17{,}700} \qquad \text{(A-16)}$$

in the regenerative case. The stone flow rate is given by

$$N_G = \frac{41.3\,Y\varepsilon_s}{1 - 88\,X_c} \qquad \text{(A-17)}$$

With regenerator temperature control by stone circulation, ($N_{MR} = 0$), and

$$N_G = N_{G\ MIN} = \frac{Y\varepsilon_s}{0.045} \qquad \text{(A-18)}$$

With regenerator temperature control by make-up stone, ($N_G = N_{G\ MIN}$), or regenerator heat transfer surface. For the once-through operation

$$1-\varepsilon_T = \frac{\varepsilon_s Y\,(10{,}630) + 9651\,Y - 1.134 \times 10^4\,Y\varepsilon_s + 177}{17{,}700} \qquad \text{(A-19)}$$

which is independent of X_c because of the recycling of the combustion energy of the deposited carbon from the sulfate generator to the gasifier.

The disadvantage of using the stone circulation regenerator temperature control method is evident from Figure A-6 as X_c increases above 0.006. Although Esso (England) has normally obtained lower values of X_c less than 0.006 at 20% of stoichiometric air/fuel ratios or higher, high values of X_c may be obtained at lower air/fuel ratios, calling for extremely high stone circulation rates and low thermal efficiencies.

GASIFIER PRODUCT ADIABATIC FLAME TEMPERATURE

An energy balance around the gasifier and combustion region of the boiler, assuming adiabatic flame conditions, yields

$$\varepsilon_T \Delta H_C = F_{cp}\,\bar{C}'_p\,(T_F - 77) + F_s\,\bar{C}_{ps}\,(T_F - T_s) + F_s\lambda_v\ . \qquad \text{(A-20)}$$

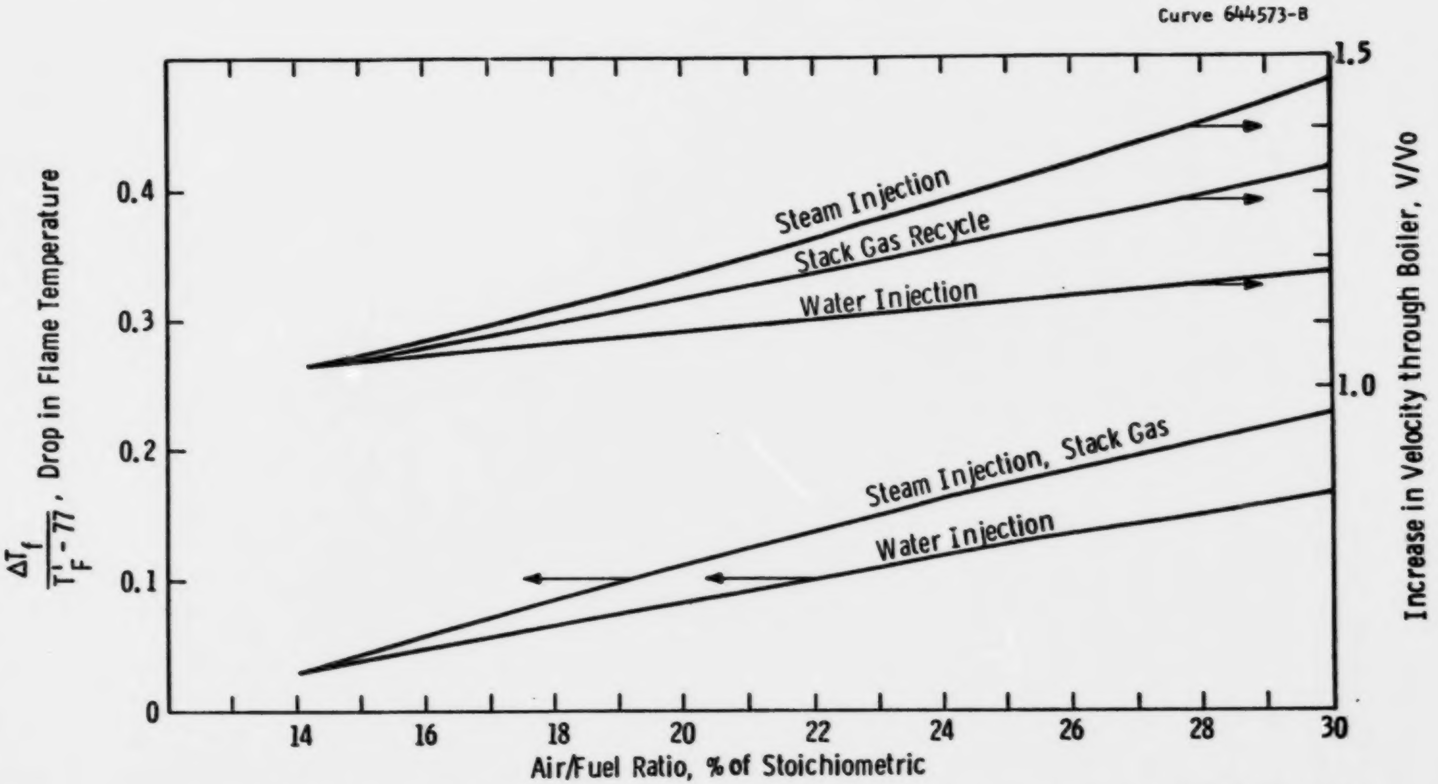
Where ε_T is the thermal efficiency of the gasifier, excluding the lost heat of vaporization (as is shown in Figure A-6) ΔH_C is the fuel heat of combustion, 17,700 Btu/lb; F_{cp} is the mass flow rate of the combustion products in lb/lb fuel; $\bar{C}_p'$ is the mean heat capacity of the combustion products, ∿0.30 Btu/lb°F; T_F is the adiabatic flame temperature; F_s is the mass flow rate of the temperature control stream in lb/lb fuel; $\bar{C}_{ps}$ is the mean heat capacity of the temperature control medium; T_s is the entrance temperature of the temperature control stream into the gasifier; and λ_v is the latent heat of vaporization of the temperature control medium. If the fuel oil were burned directly, without gasification, then the energy balance would yield, approximately,

$$\Delta H_C = F_{cp}\,\bar{C}_p\,(T_F' - 77) \tag{A-21}$$

where T_F' is the adiabatic flame temperature without gasification. Thus, from equations A-20 and A-22

$$\frac{T_F' - T_F}{T_F' - 77} = 1 - \frac{\varepsilon_T - \dfrac{F_s \lambda_v}{\Delta H_s} + \dfrac{F_s\,\bar{C}_{ps}\,(T_s - 77)}{\Delta H_C}}{1 + \dfrac{F_s\,\bar{C}_{ps}}{F_{cp}\,C_p}} \tag{A-22}$$

to give the reduction in gasifier product flame temperature as a function of the temperature control scheme and the air/fuel ratio. Equation A-22 is shown in Figure A-7 for stack gas recycle, steam injection, and water injection temperature control schemes with 95% sulfur removal of a 3 wt % sulfur fuel oil, 10% excess combustion air, and $X_c = 0.003$. The curves shown represent the average behavior of regenerative and once-through operations with the various regenerator control methods, all these cases giving similar results in equation A-22. The relative velocity of the combustion products through the boiler, $\bar{V}/\bar{V}_o$, compared to the velocity, $\bar{V}_o$, without gasification is also shown in Figure A-7 as a function of the air/fuel ratio and the gasifier temperature control



Basis: 3 wt % S Fuel, 95% Removal, $X_c = 0.003$, 10% Excess Combustion Air in Boiler; Lines Represent Average Behavior for Regenerative and Once-through Operation, and the Three Regenerator Temp. Control Schemes.

Figure A-7—Boiler Derate Factors versus Air/Fuel Ratio and Gasifier Temperature Control Scheme method. The figure indicates that a considerable decrease in the flame temperature can be expected if high air/fuel ratios are used (about a 400 to 900°F drop expected at an air/fuel ratio of 30% of stoichiometric), while a great increase in the velocity through the boiler may be simultaneously obtained at high air/fuel ratios ($\overline{V}_o \approx 50$ ft/sec; $\overline{V} \approx 75$ ft/sec with steam injection at an air fuel ratio of 30% of stoichiometric and 60 ft/sec with water injection at an air/fuel ratio of 30% of stoichiometric). The combined effects of these two variables on the boiler performance, plus the additional factor of flame luminosity, are not easily evaluated, but Figure A-7 indicates the desire to operate the gasifier at as low an air/fuel ratio as is possible to maintain high desulfurization and efficient overall operation. Some of these aspects of boiler performance may be investigated by means of the Esso (England) continuous gasification/desulfurization pilot plant.

APPENDIX B

SPACE REQUIREMENTS AND GASIFIER START-UP

PIPE SIZES FOR CARRYING GASIFIER PRODUCT FUEL

The pipe diameter required to carry the 1600°F fuel gas from the gasifier to the boiler is set by the capacity of the gasifier, the air/fuel ratio with which the gasifier is operated, the method of temperature control in the gasifier, plant layout, and the allowable pressure drop over the fuel pipe length. An estimate of the fuel pipe diameter is made in terms of these factors.

The hot gas heating value of the fuel may vary from 250 to 500 Btu/scf depending on the air/fuel ratio and the temperature control scheme. The high value of 500 Btu/scf corresponds to temperature control by operation at the minimum air/fuel ratio (∿ 14% of stoichiometric). The lower value of 250 Btu/scf corresponds to an air/fuel ratio of about 20% of stoichiometric and temperature control by water injection, or stack gas recycle.

Assumptions made in evaluating fuel pipe diameters are that the plant heat rate is 10^4 Btu/kWh, the fuel pipe length is 50 ft, and the pressure drop across the pipe length to be tolerated is 2 inches of H_2O. The pressure drop formula is

$$\Delta P = \frac{32\ f\ L\ Q^2\ \rho}{\pi^2\ g_c\ D^5} \quad ,$$

where f is the fanning friction factor, L is the fuel pipe length, Q is the volumetric flow rate of the fuel gas, ρ is the fuel gas density, and D is the inner pipe diameter.

Curve 647030-A

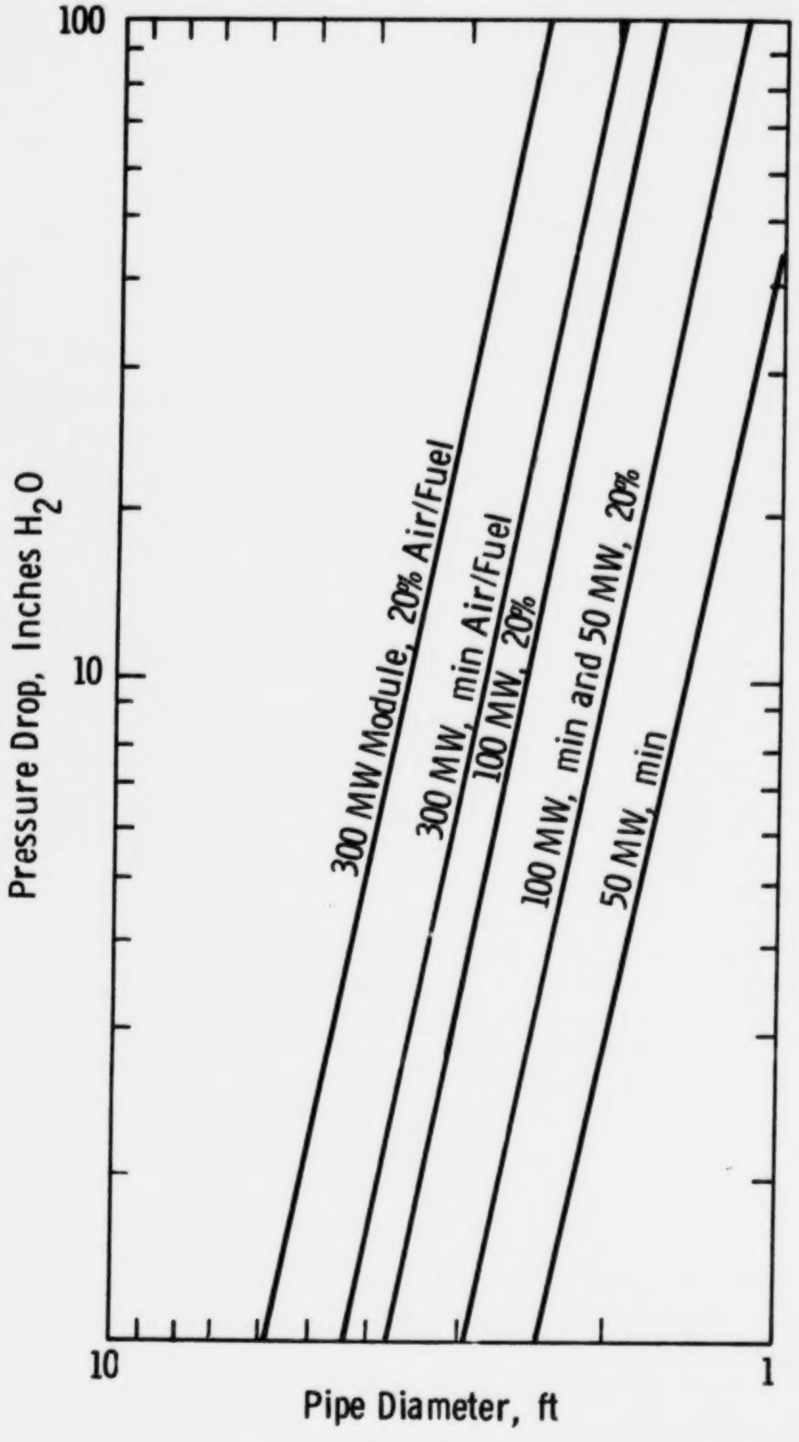


Figure B-1—Fuel Gas Transport Pressure Drop

Curve 646761-A

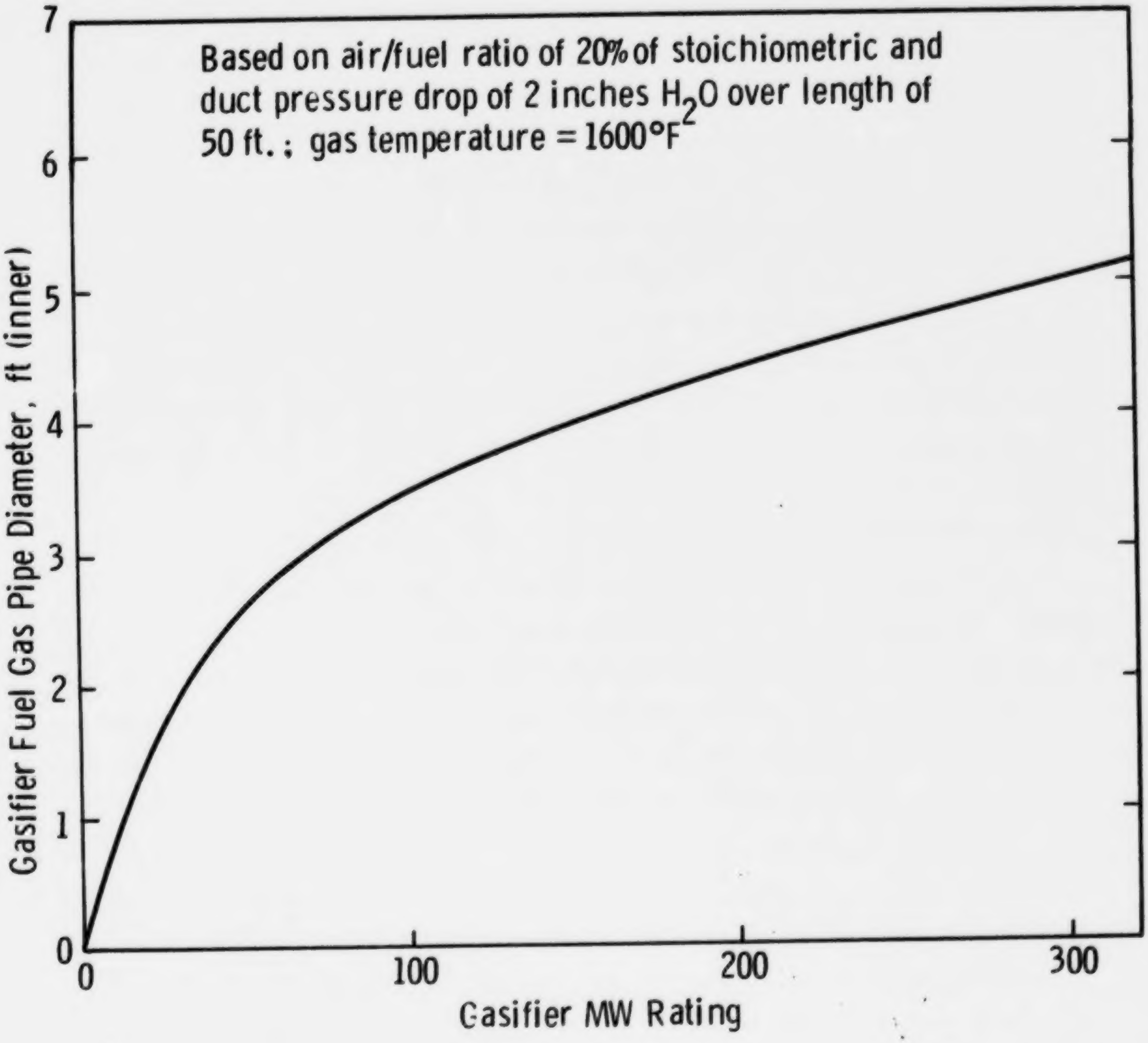


Figure B-2—Hot Fuel Gas Pipe Diameter Estimate

Evaluation of this relationship for gasifier capacities of 50, 100, and 300 MW with operation at the minimum air/fuel ratio and at 20% of stoichiometric gives the results shown in Figure B-1. The pressure drop increases rapidly with a decrease in the fuel pipe diameter and is sensitive to the air/fuel ratio. With 2 inches H_2O pressure drop as a basis, Figure B-2 indicates the fuel pipe diameter as a function of the gasifier capacity for an air/fuel ratio of 20% of stoichiometric. The inner diameter goes from 3 feet for a 100 MW gasifier to 5 feet for a 300 MW gasifier.

The fuel pipe will consist of about 6 inches of refractory lining, 4 to 5 inches of insulation, and an outer layer of carbon steel. This will make the outer fuel pipe diameter about 2 feet greater than the inner diameter. The pipe weight will be about 1500 lb per ft of length.

SPACE REQUIREMENTS

Gasifier, regenerator, and sulfate generator vessel dimensions have been evaluated for the factors of air/fuel ratio, temperature control scheme, fluidizing velocity, and gasifier capacity. Figure B-3 shows the cross-sectional area of the gasifier, regenerator, and sulfate generator vessels plotted against the gasifier MW rating for the case of an air/fuel ratio of 17% of stoichiometric and a fluidizing velocity of 8 ft/sec.

The vessels will be about 15 to 20 feet tall. With either the once-through system or the regenerative system, the cross-sectional areas of the vessels are less than the cross-sectional area of coal- or oil-fired boiler of the same MW rating. The atmospheric pressure oil gasification concept is much less space consuming than stack gas cleaning systems.

GASIFIER START-UP

A variety of gasifier start-up procedures can be applied to raise the temperature of the gasifier vessel and limestone load from its

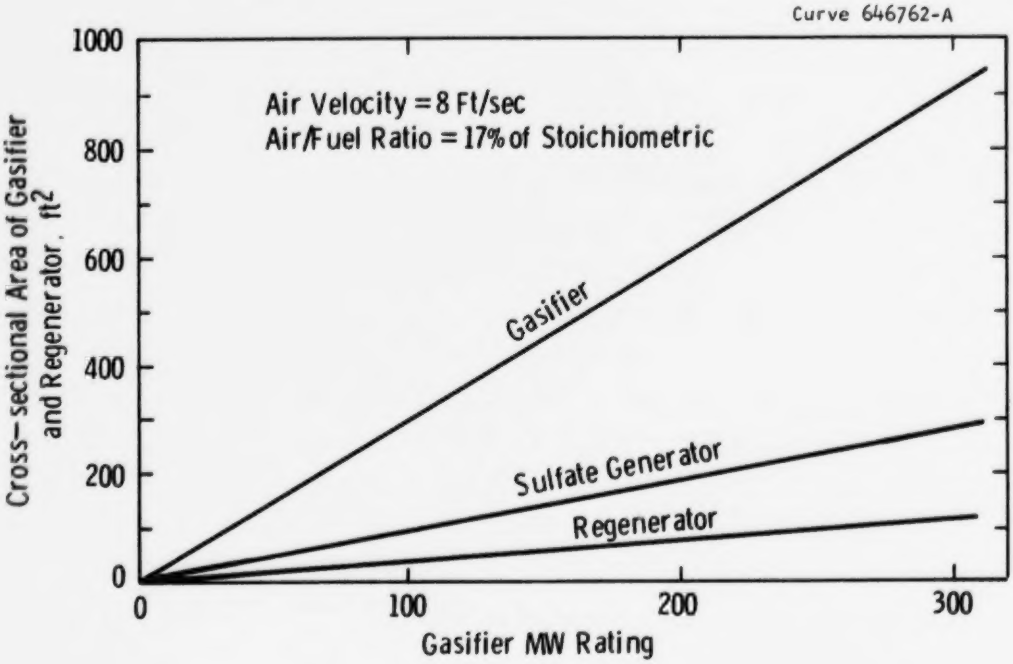


Figure B-3—Reactor Cross-Section

initial cold state up to 1600°F. The rate of the vessel temperature increase must be slow enough to avoid damage to the vessel due to thermal expansion stresses. The start-up scheme should provide the gasifier with a load of 1600°F lime (CaO) which has not lost its activity due to exposure to extremely high temperatures. The start-up procedure examined heats the bed by using a burner fired with a clean auxiliary fuel while fluidizing the amount of limestone needed for operation at about twice the minimum fluidization velocity (about 2 ft/sec with 1000 μ diameter particles). The burner (or burners) would be located in the gasifier wall near the air distributor plate and would yield an average air inlet temperature of 1800°F. The heat transfer characteristics in a bed of 1000 μ particles are such that most of the heat provided to the bed will be utilized to calcine the limestone with little increase in the gasifier temperature once the bed has reached reaction temperature (∿ 1200°F). The calcination period will be followed by a quick heat-up of the lime bed, a slower heat-up of the vessel which is in contact with the fluidized bed, and a very slow heat-up of the vessel walls above the level of the bed. This behavior is demonstrated in Figure B-4 for a 300 MW gasifier. Once the lime bed has reached 1500°F, high-sulfur residual oil injection may begin (after about 12 hours); and the bed may be quickly brought up to full capacity operation. This start-up procedure, which should ensure long refractory life and maintenance of lime activity by avoiding extremely high temperatures, would require about 40 x 10^6 Btu/hr.

Maintenance of the bed temperature at 1000°F during periods of shutdown will cost about 0.7¢/MM Btu based on a clean auxiliary fuel cost of $1.00/MM Btu and a 50% capacity factor. This would permit start-up of the bed in about 1.5 hours.

MODULAR GASIFIER DESIGNS

It appears that the possible advantages of using a large number of gasifier modules in a power plant are limited by economic considerations. If, for example, a 600 MW power plant were equipped with

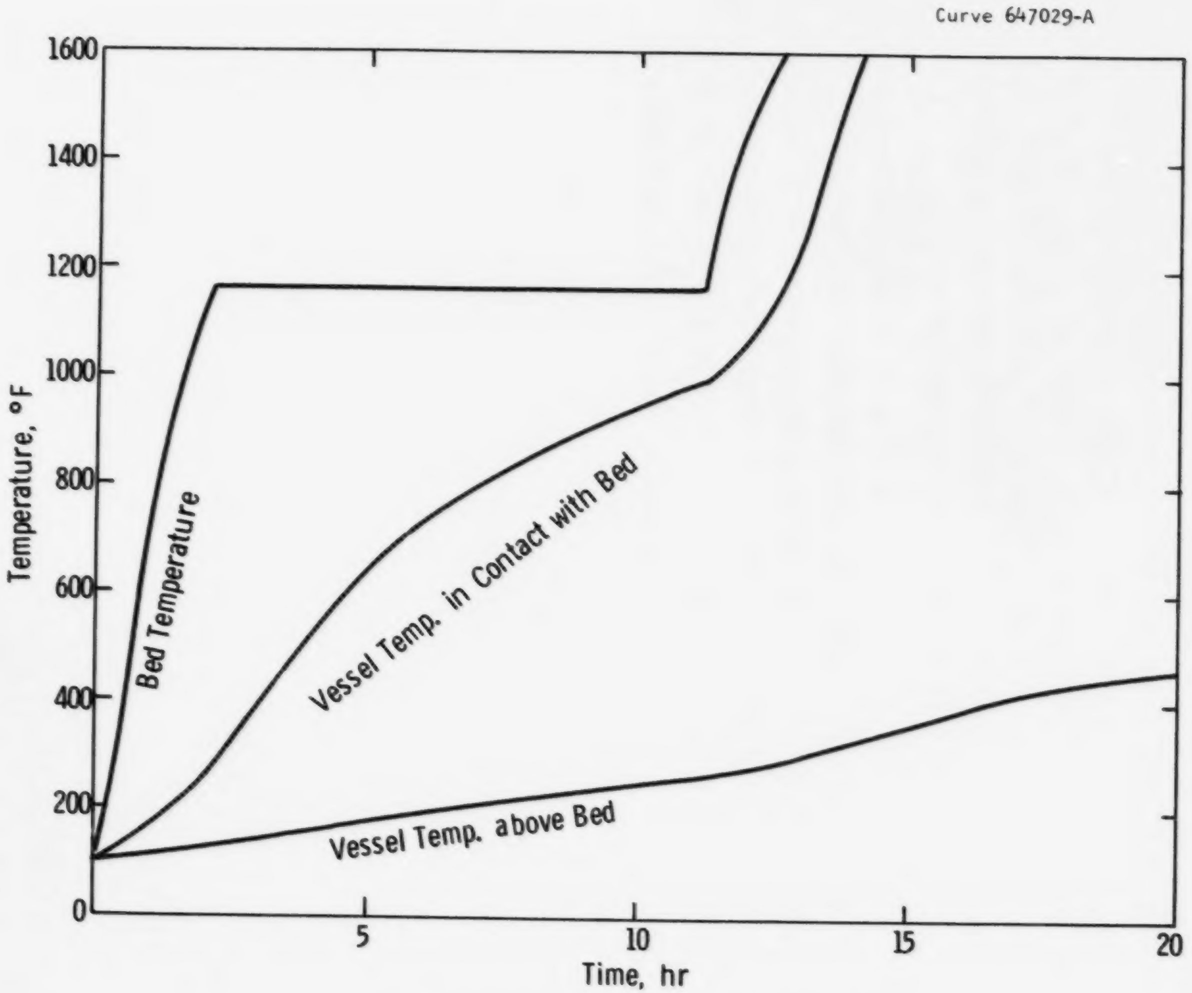


Figure B-4—Start-up of a 300 MW Gasifier

20 gasifier modules rather than the two modules previously considered for such a plant, piping sizes would be reduced from about 5 ft inner diameter to about 2 ft inner diameter and the gasifier diameter would be reduced from about 35 ft to about 11 ft. Though this size reduction to 30 MW modules may permit prefabrication of the gasifier vessels, the large number of modules would occupy more space than the two 300 MW modules, and the pipe size reduction would not be great.

If a 0.7 factor is used to scale the cost of the gasifier modules (including the regenerator or sulfate generator and the solid circulation equipment), then the increase in the number of modules from two 300 MW modules to twenty 30 MW modules would increase the total system cost from about \$20/kw to about \$26/kw, a 30% increase in cost.

These preliminary considerations indicate that the number of modules should be limited to no more than four.

APPENDIX C

EVALUATION OF OIL GASIFICATION PROCESS COSTS

Capital costs and fuel adder costs have been investigated for 600 MW regenerative and once-through operations to show the effects of the various operating and design variables.

REGENERATIVE AND ONCE-THROUGH CAPITAL COSTS

- Basis
  - 600 MW capacity retrofit
  - Plant heat rate of 10^4 Btu/kWh
  - Gasifier temperature control with minimum air/fuel ratio
  - Fuel oil low heating value of 17,700 Btu/lb.

The costs of the components of the regenerative system are estimated as follows,based on material costs for stainless steel shells and expected distributor plate costs. Vessel costs are estimated to be given by

- Gasifier Vessel Cost $= 3.34 \frac{(MW)}{V\epsilon} + 9.96 \left(\frac{MW}{V\epsilon}\right)^{1/2}$ (C-1)*

- Regenerator Vessel Cost $= 273.0 \left\{ \frac{X_c N_G}{12} + 0.053\, Y\epsilon_s \right\} \frac{(MW)}{V\epsilon}$

$$+ 90.0 \left[\left\{ \frac{X_c N_G}{12} + 0.053\, Y\epsilon_s \right\} \frac{(MW)}{V\epsilon} \right]^{1/2} . \qquad \text{(C-2)}$$

At 600 MW capacity limestone handling, disposal, and circulation costs are computed as

$$2.59 \times 10^7 \left[\frac{CaO}{S}\right] \frac{Y}{\epsilon} + 8.81 \times 10^5 \frac{NG}{\epsilon} \quad . \qquad \text{(C-3)}$$

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*See nomenclature listings for explanation of symbols, p. C-8.

Uninstalled fan costs are taken at \$0.33/kw, burners at \$1.0/kw, particulate cleaning equipment at \$2.66/kw, and boiler modifications at \$1.50/kw. The capital cost, installed, of the sulfur recovery plant, C_s, is a function of the SO_2 concentration from the regenerator and the net tons of sulfur produced per day, or

$$C_s = \text{function}\ (X_{SO_2}\ , \text{NTD})\ . \qquad \text{(C-4)}$$

With a factor of 50% of the purchased equipment cost taken for installation, and a factor of 30% of the installed equipment cost taken for engineering, contingency, and construction, the total investment per kw is found to be

$$\begin{aligned}\$/\text{kw} = {} & (6.51 + 44.5\ X_c N_G + 28.3\ Y\varepsilon_s)\ \frac{1}{V\varepsilon} \\ & + (0.793 + 2.07\ \{X_c N_G\}^{1/2} + 1.65\ \ Y\varepsilon_s{}^{1/2})\ (\frac{1}{V\varepsilon})^{1/2} \qquad \text{(C-5)} \\ & + 84.1\ [\frac{CaO}{S}]\ \frac{Y}{\varepsilon} + 2.86\ \frac{N_G}{\varepsilon} + 10.71 + \frac{C_s}{600{,}000}\ .\end{aligned}$$

In this equation X_c is the wt fraction of carbon on the gasifier stone, N_G is the flow rate of sulfided lime from the gasifier to the regenerator, in lb/lb fuel, given by

$$N_G = \frac{41.3\ Y\varepsilon_s - 14.5\ N_{MR}}{1 - 88\ X_c}\quad , \qquad \text{(C-6)}$$

where N_{MR} is the rate at which make-up limestone is added to the regenerator, in lb/lb fuel. Y is the wt fraction of sulfur in the fuel oil and ε_s is the sulfur removal efficiency. V is the superficial velocity in ft/sec, assumed to be the same in the gasifier and regenerator, and ε is the overall system efficiency, including heat lost to the surroundings and energy lost in supplying the system power requirement. The factor [CaO/S] is the molar ratio of lime to sulfur fed in the make-up limestone stream and is a function of the sulfur removal efficiency estimated from the Esso (England) data. C_s is the installed cost of a sulfur recovery plant and is a function of X_{SO_2} and NTD where

$$X_{SO_2} = \frac{Y\varepsilon_s}{12.7\ X_c N_G + 7.39\ Y\varepsilon_s} \tag{C-7}$$

and

$$N_{TD} = 6.78\ \frac{Y\varepsilon_s}{\varepsilon}\ (MW)\quad , \tag{C-8}$$

where (MW) is the capacity of the plant in MW.

Similar considerations for a 600 MW once-through system give

$$\begin{aligned} \$/kw &= (6.51 + 151.5\ X_c N_G + 113.7\ Y\varepsilon_s)\ \frac{1}{V\varepsilon} \\ &+ (0.793 + 3.93\ \{X_c N_G\}^{1/2} + 3.41\ \{Y\varepsilon_s\}^{1/2})\ (\frac{1}{V\varepsilon})^{1/2} \\ &+ 48.4\ \frac{Y\varepsilon_s}{U} + 11.42\quad , \end{aligned} \tag{C-9}$$

where U is the lime utilization in the gasifier and N_G is given by

$$N_G = 1.80\ \frac{Y\varepsilon_s}{U}\quad . \tag{C-10}$$

Equations C-5 and C-9, for regenerative and once-through operations, are shown in Figure C-1 as a function of ε_s with V = 8 ft/sec, Y = 0.03, U = 0.35, 1% heat losses to the surrounding, 1% power requirement, regenerator temperature control by stone circulation ($N_{MR} = 0$) and by make-up stone ($N_G = N_{G\ MIN}$), with X_c as a parameter. Figure C-1 indicates that the economics of the regenerative operation with regenerator temperature control by stone circulation is very sensitive to X_c and ε_s (because of the high stone circulation required at high X_c and the high expense of sulfur recovery at high X_c), while the cost with regenerator temperature control when make-up stone is fed to the regenerator is relatively insensitive to X_c. The capital investment in the once-through operation is less than the regenerative operation investment and is insensitive to the sulfur removal efficiency and nearly independent of X_c.

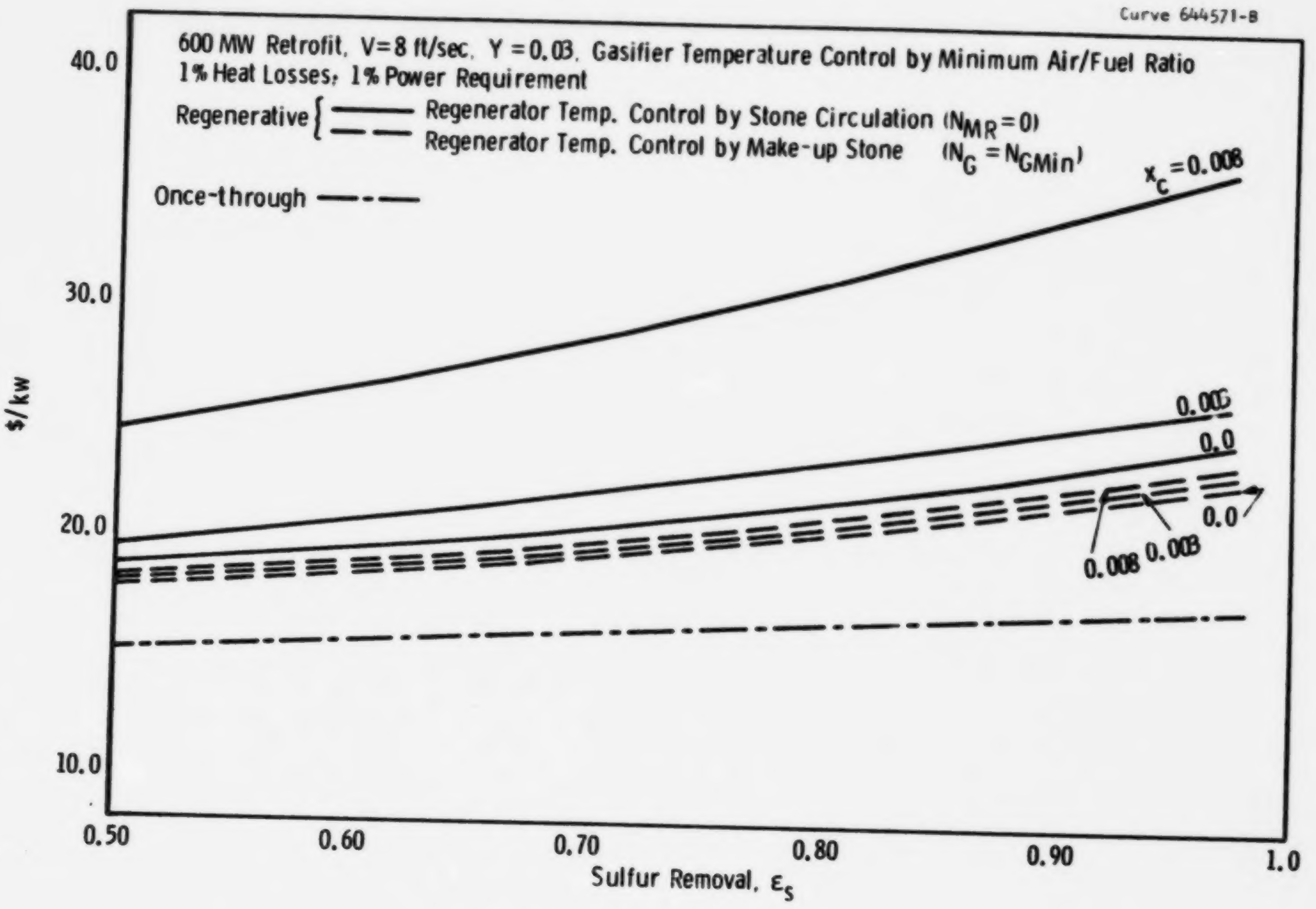

Figure C-1—Capital Cost versus Sulfur Removal

FUEL ADDER COSTS

With capital charges of 14% of the total investment, maintenance at 5%, and overhead at 4%, the fuel adder in $¢/10^6$ Btu for a 600 MW regenerative operation is given by

$$2.625 \times 10^{-1} \frac{I}{L} + 17.0 \frac{1-\varepsilon}{\varepsilon} C_F + \left[\frac{CaO}{S}\right] \frac{Y}{\varepsilon} 8.96 C_L + 2.825 \frac{Y\varepsilon_s}{\varepsilon} (OC)_s \; . \qquad \text{(C-11)}$$

I is the total investment in \$/kw, L is the plant load factor, C_F is the cost of residual fuel oil in \$/bbl, C_L is the purchase price of limestone plus the disposal cost of spent lime in \$/ton of limestone, and $(OC)_s$ is the operating cost of the sulfur recovery plant in \$/net ton of sulfur, which is a function of X_{SO_2} and NTD. Similarly, for a 600 MW once-through operation,

$$¢/10^6 \text{ Btu} = 2.625 \times 10^{-1} \frac{I}{L} + 8.93 \frac{Y\varepsilon_s}{\varepsilon U} C_L + 17.0 \frac{1-\varepsilon}{\varepsilon} C_F \, . \qquad \text{(C-12)}$$

Equations C-11 and C-12 are shown in Figures C-2 and C-3 as a function of the sulfur removal efficiency. Regenerative and once-through operation fuel adder costs are shown for fuel costs of \$2 and \$5/bbl in the two figures. For the regenerative operation, load factors of 40% and 80% are considered and C_L = \$8, \$4, and \$2/ton are shown. The additional parameter of lime utilization, U, is considered in the once-through operation with U = 0.35 and 0.25. For the parameter values considered, the once-through operation fuel adder cost is lower than the regenerative operation fuel adder cost except at the lower utilization and higher limestone costs. The load factor has a large effect on the fuel adder cost while, for the fairly high efficiency situation considered, the fuel cost has a very small effect on the fuel adder. Should heat losses or power requirements increase 1 or 2%, the fuel cost factor will increase in importance, as is seen from the factor $(1-\varepsilon)/\varepsilon$ in equations C-11 and C-12. Because of the high investment in the regenerative operation, compared

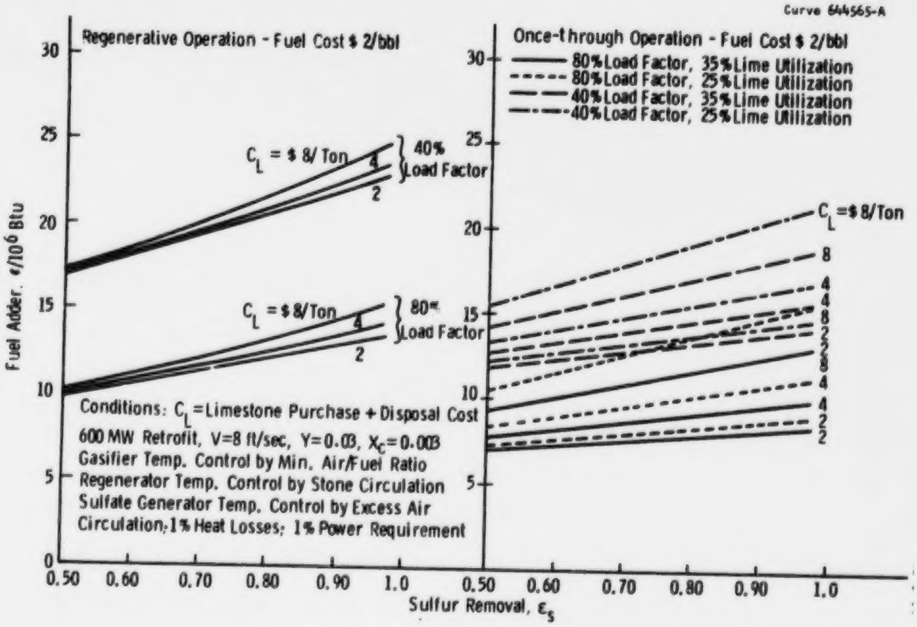

Figure C-2—Fuel Adder Cost versus Sulfur Removal

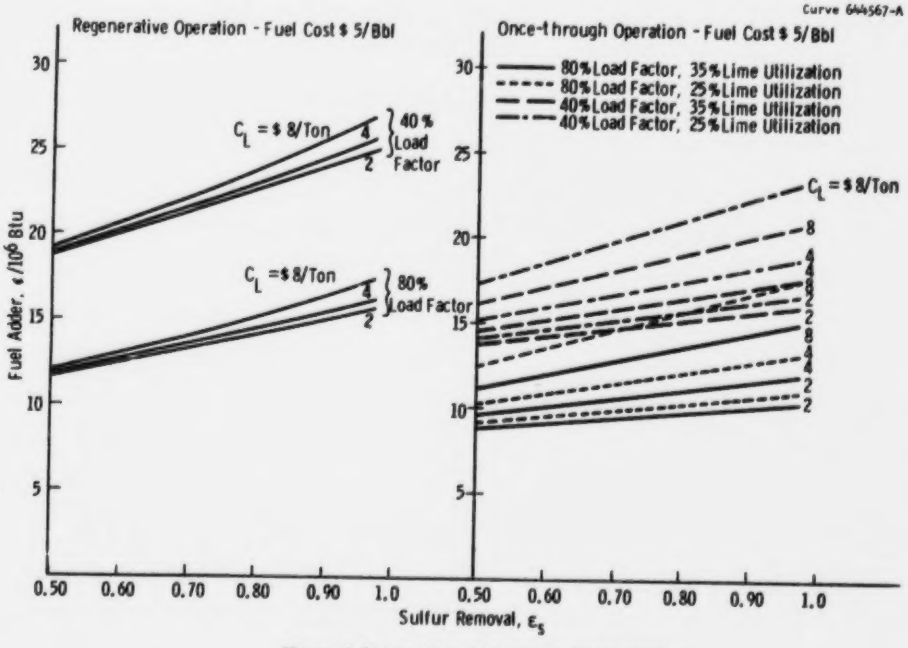

Figure C-3—Fuel Adder Cost versus Sulfur Removal to the once-through operation, the load factor affects the regenerative fuel adder cost more than it does the once-through fuel adder cost. Thus, at load factors as low as 40%, the once-through process has a 3 to 8¢/10^6 Btu lower fuel adder cost than the regenerative operation, depending on the limestone purchase and disposal cost.

The economic factors considered here lead to a greater understanding of the feasibility of the atmospheric-pressure oil gasification concepts and provides preliminary information useful in more detailed design phases.

NOMENCLATURE

C_F - The cost of residual fuel oil, in \$/bbl.

C_L - The purchase cost of limestone plus the disposal cost, in \$/ton.

$\bar{C}_p$ - The mean heat capacity of the gasifier product stream, excluding the contribution of a temperature control component, in Btu/lb °F.

C'_p - The mean heat capacity of the products of combustion in the boiler.

C_s - The capital cost associated with a sulfur recovery plant recovering elemental sulfur, in \$.

$\bar{C}_{ps}$ - The mean heat capacity of the temperature control medium, stack gas, steam, or water, in Btu/lb °F.

[CaO/S] - The molar feed ratio of CaO to sulfur in the regenerative operation.

F_p - The mass flow rate of product from the gasifier, in lb/hr/lb fuel/hr.

F_s - The mass flow rate of the temperature control stream in the gasifier, in lb/hr/lb fuel/hr.

F_{cp} - The mass flow rate of the products of combustion in the boiler, in lb/lb fuel.

I - The total investment in the gasification/desulfurization system, in \$/kw.

K - K_2/K_1.

K_1 - $M_{CO} + M_{CO_2}$, the molar flow rate of oxidized carbon from the gasifier, in moles/lb fuel.

K_2 - $M_{O_2}(I)$.

K_E - The equilibrium constant for the reaction

$$CO_2 + H_2 \rightleftarrows CO + H_2O \ .$$

L - The plant load factor.

MW - The plant generating capacity, in MW.

M_i - The molar flow rate of species i, moles/lb fuel.

$M_{O_2}(I)$ - The molar feed rate of O_2 to the gasifier, in moles/lb fuel.

N_G - The mass flow rate of lime from the gasifier to the regenerator or sulfate generator, in lb/lb fuel.

$N_{G\ MIN}$ - The minimum allowable mass flow rate of lime based on the maximum allowable lime utilization in the gasifier.

N_{MR} - The limestone make-up rate to the regenerator, in lb/lb fuel.

NTD - The net ton of sulfur recovered per day in the sulfur recovery plant.

$(OC)_s$ - The operating cost associated with the sulfur recovery plant, in $/net ton of sulfur.

T_F - The adiabatic flame temperature, °F.

T'_F - The adiabatic flame temperature obtained without gasification of the oil prior to burning, °F.

T_s - The temperature of the temperature control stream entering the gasifier, °F.

U - The CaO utilization in the gasifier; mole fraction.

V - The superficial velocity, ft/sec.

$\bar{V}$ - The average velocity of combustion products through the boiler, ft/sec.

$\bar{V}_o$ - The average velocity of combustion products through the boiler without gasification, ft/sec.

X_c - The mass fraction of carbon deposited on the gasifier lime.

X_{SO_2} - The mole fraction of SO_2 in the regenerator product.

Y - The mass fraction of sulfur in the fuel oil.

$\Delta H_{control}$ - The net heat production in the gasifier with no energy losses, Btu/lb fuel.

ΔH_C - The heat of combustion of the fuel oil.

ΔH_G - The total heat production rate in the gasifier due to combustion, Btu/lb fuel.

ΔT_p - The change in gasifier product temperature, = 1523°F.

ε - The overall system efficiency including power requirements.

ε_s - The sulfur removal efficiency.

ε_T - The gasifier thermal efficiency.

λ_v - The heat of vaporization of the temperature control stream, in Btu/lb.

APPENDIX D

OIL GASIFIER TURNDOWN -- VARIABLE AIR/FUEL RATIO TO MAINTAIN FLUIDIZATION DURING TURNDOWN

The fluidization velocity in the oil gasifier may be maintained above the minimum fluidization velocity during gasifier turndown by increasing the air/fuel ratio as the fuel rate is decreased. Figure D-1 shows that the heat generated in the gasifier per lb fuel oil increases linearly with the air/fuel ratio in the range of air/fuel ratios studied by Esso (England). The data shown have been extrapolated to lower air/fuel ratios and applied to an energy balance to give the relationships shown in Appendix A. In this appendix the gasifier temperature control stream flow rate per lb of fuel oil is shown as a function of the air/fuel ratio. Stack gas recycle, steam injection, and water injection are considered. Figure D-2 shows the gasifier thermal efficiency as a function of the air/fuel ratio for the three temperature control methods.

To evaluate gasifier turndown, the following assumptions are made:

- The gasifier can be operated at the minimum air/fuel ratio of ∿ 14% of stoichiometric when operating at full capacity. The control temperature is 1600°F at all air/fuel ratios.
- The ratio of the fluidization velocity at turndown to the fluidization velocity at full capacity, V_f/V_f°, is given by

$$V_f/V_f^\circ = \frac{M_f/M^\circ{}_f[0.198(A/F) + 0.0144(M_T/M_f)}{[0.198(A/F)^\circ + 0.0144(M_T/M_f)^\circ} \quad ,$$

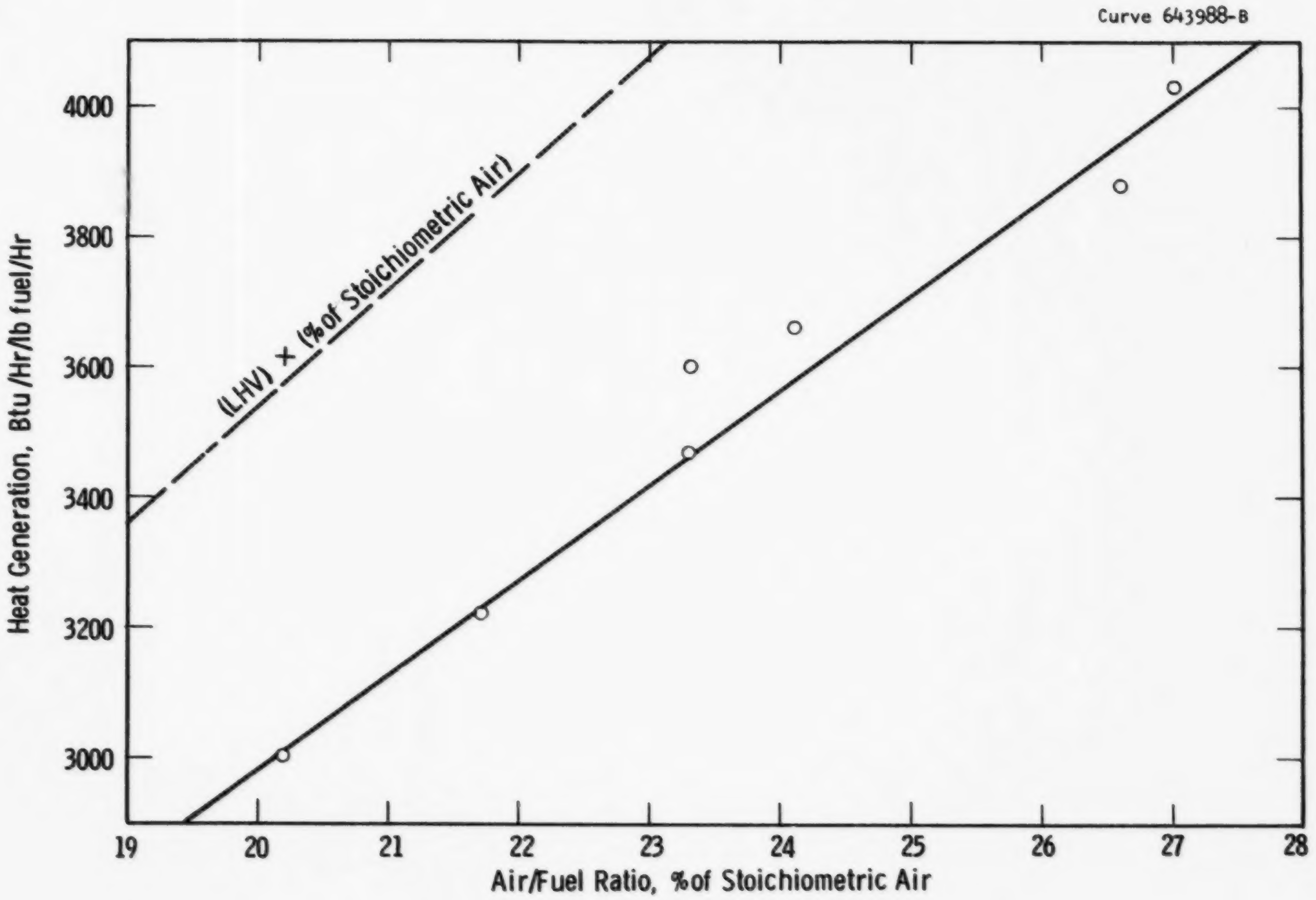

Figure D-1—Gasification-Desulfurization Heat Generation Rate

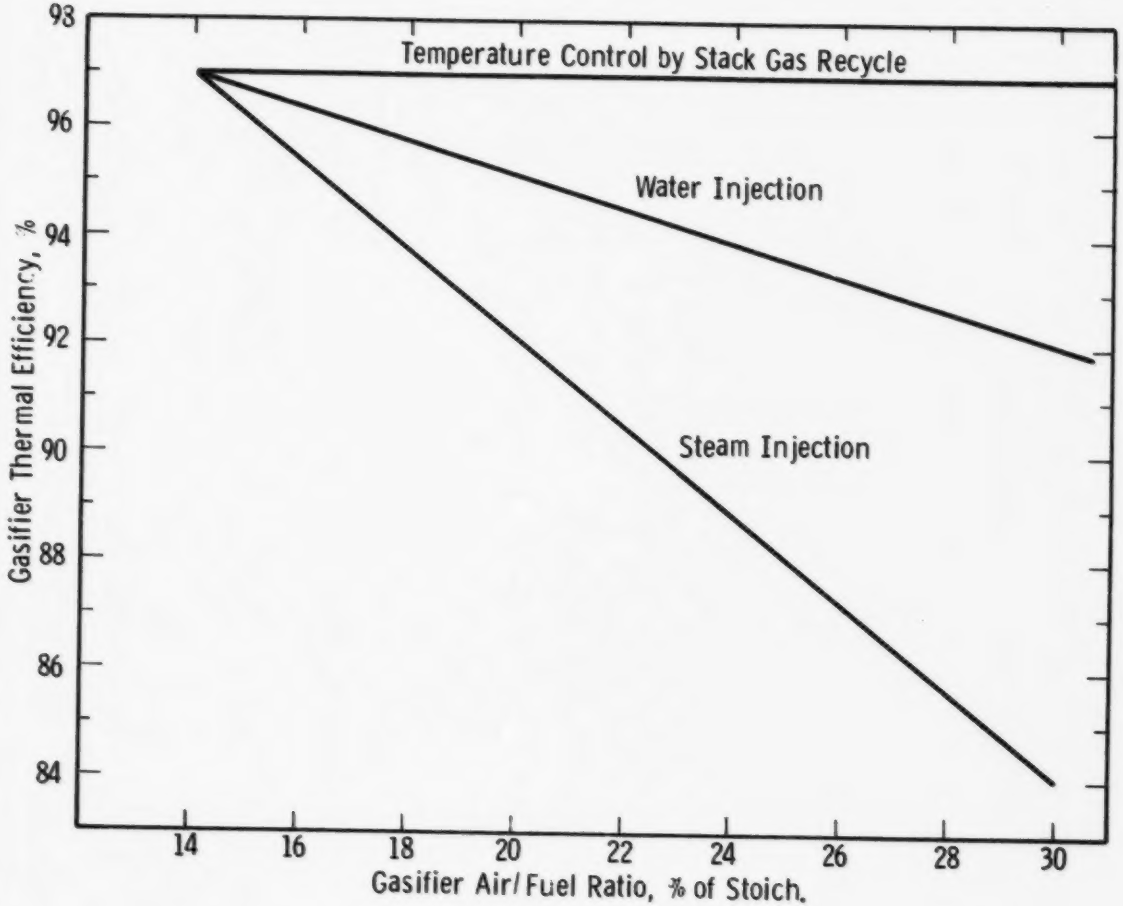

Figure D-2—Gasifier Thermal Efficiency

where

M_f/M_f° = fraction of full capacity or the fuel ratio at turndown;

(A/F) = the air/fuel ratio, fraction of stoichiometric;

M_T/M_f = the mass ratio of the temperature control stream to the fuel rate (shown in Appendix A);

$(A/F)^\circ$ = the air/fuel ratio at full capacity (= 0.14 from assumption 1);

$(M_T/M_f)^\circ$ = temperature control stream/fuel ratio at full capacity (= 0 from assumption 1 and Appendix A).

- The ratio of the hot fuel gas heating value at turndown to the heating value at full capacity is given by

$$HV/HV^\circ = (V_f^\circ/V_f)(M_f/M_f^\circ) \quad .$$

- An arbitrary air/fuel ratio policy must be chosen to relate the air/fuel ratio to the turndown. The following policy is evaluated here:

$$(A/F) = (A/F)^\circ \frac{1}{\dfrac{1 - (1-M_f/M_f^\circ)}{2}} \quad .$$

Figure D-3 shows the fluidization velocity ratio as a function of the fraction of full capacity down to 0.25 (4/1 turndown). Also shown is the fuel gas heating value ratio. The heating value of the fuel gas at full capacity may be about 450 Btu/scf. The temperature control schemes of stack gas recycle and water injection are considered. The velocity ratio for a policy of constant air/fuel ratio during turndown is also indicated in the figure for comparison.

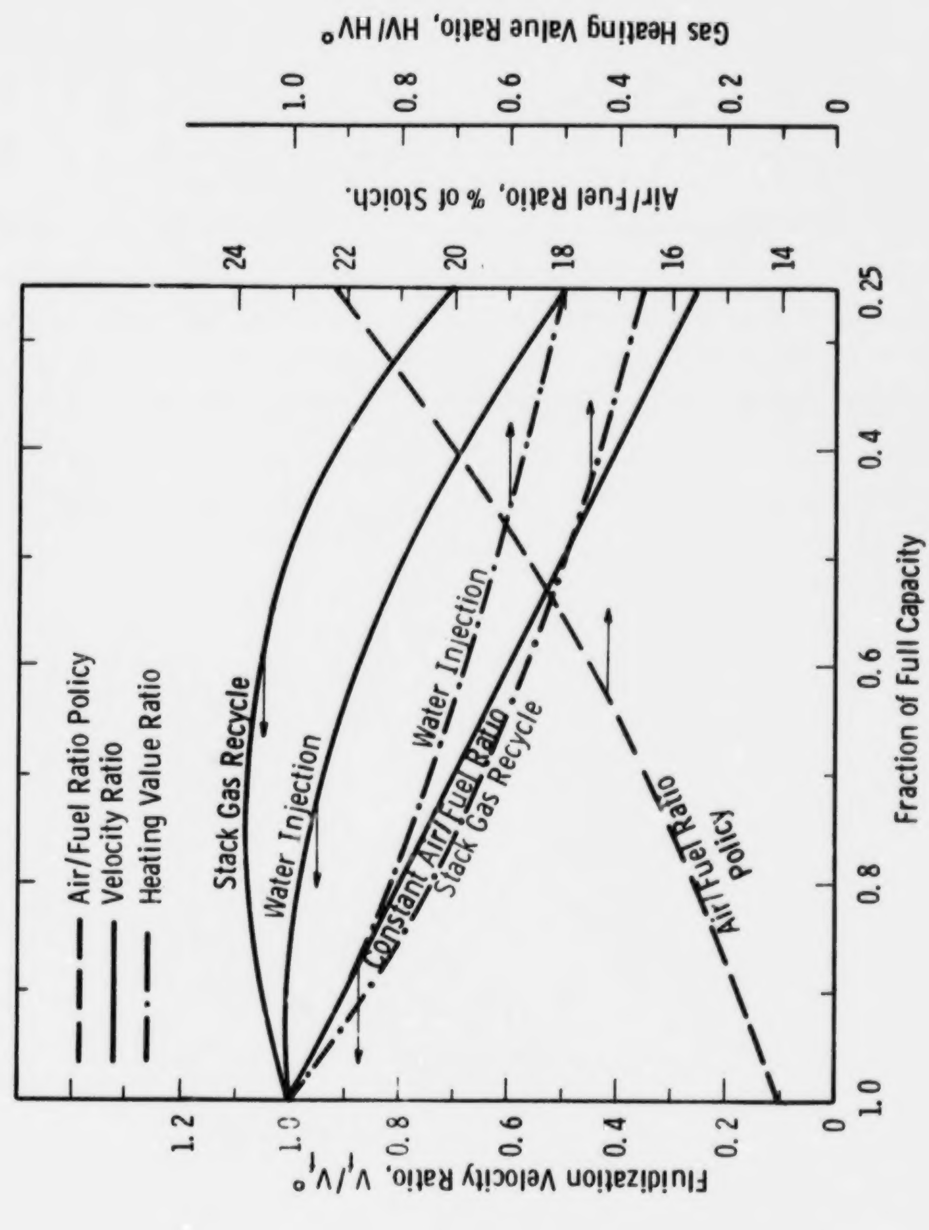

Figure D-3—Gasifier Turndown

In testing an air/fuel ratio policy the gasifier thermal efficiency and the fuel gas heating value must be considered. For example, the policy

$$A/F = (A/F)^{\circ}(M_f^{\circ}/M_f)$$

gives a heating value of only about 70 Btu/scf at 25% of full capacity which may provide poor combustion properties. It must be realized that the gasifier may not be able to operate at an air/fuel ratio as low as 14% of stoichiometric, and operating at higher air/fuel ratios at full capacity (say 20%) makes this method of turndown less feasible.

For the case studied in the last figure, turndown to 25% appears realistic. A commercial system may require only a 50% turndown capability because of modular gasifier design.

APPENDIX E

PRESENTATION DOCUMENT[*] -- CLEAN BOILER FUEL BY FLUIDIZED BED GASIFICATION/DESULFURIZATION

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[*] This document was prepared January 1972. The document is presented without change. Significant changes have occurred since the writing of this document, for example, fuel prices, schedule, spent solids processing, pilot plant data. Thus, the information presented does not necessarily represent the current program status.

CLEAN BOILER FUEL

BY

FLUIDIZED BED GASIFICATION/DESULFURIZATION

A program carried out by

Westinghouse Electric Corporation
Pittsburgh, Pennsylvania

for the

Office of Research and Monitoring
Environmental Protection Agency
Research Triangle Park,
North Carolina 27711

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I. INTRODUCTION

II. OIL FUEL SITUATION

III. BASIC PROCESS CONCEPTS

IV. EVALUATION OF FLUIDIZED BED GASIFICATION/DESULFURIZATION

- A. Conceptual Designs
- B. Assessment
- C. Experimental Development
- D. Demonstration Plant

V. PROPOSED DEMONSTRATION PLANT PROGRAM

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I. INTRODUCTION

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OIL GASIFICATION/DESULFURIZATION

Westinghouse and other forecasters predict a tremendous increase in fuel oil demand for electric utility generation in this decade. In 1980 utility fuel oil consumption is forecast to be about 2.5 times the 1970 consumption, and in 1985 about 3 times. Oil will have to make up energy deficiencies resulting from nuclear delays, the enactment of air pollution regulations which prohibit the use of high sulfur coal, and shortages of natural gas. This fuel oil will have to be of the high quality low sulfur type with inherent high costs, or processes for removing the sulfur from the oil or from the combustion gases will have to be developed and applied.

Although many fuel oil desulfurization facilities are being installed, it does not appear that adequate quantities of low sulfur oil will be available to meet future demand. Oil companies are hesitant to invest additional sums for more facilities. Sulfur dioxide stack removal process equipment is costly and smooth operation has not yet been proved.

In areas where natural gas is being curtailed utilities are experiencing difficulty burning even high quality liquid fuels in boilers that were designed specifically for gas.

Westinghouse, under the sponsorship of the Environmental Protection Agency, EPA, has carried out analyses which indicate that it may be feasible to gasify low cost high sulfur residual oil for 10-18¢/10^6 Btu. This cost compares favorably with the estimated cost differential between low and high sulfur petroleum products -- 25-50¢/10^6 Btu. In addition, gasification effectively allows poor quality resid to be burned like natural gas without incurring boiler problems.

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The Office of Research and Monitoring, ORM, of the Environmental Protection Agency, EPA, has undertaken a program to develop an atmospheric pressure fluidized bed oil gasification/desulfurization process to provide clean fuel to boilers. Under contract to ORM, Esso Petroleum Company (England) has collected basic process data. Its effort has culminated in the construction and operation of a one megawatt pilot gasifier. Utilizing Esso data Westinghouse has produced and evaluated conceptual designs of oil gasification/desulfurization systems for full scale power plants. The concept appears technically feasible; effective in reducing SO_2, NO_x, and particulate emissions; and economically competitive with alternative sulfur control options such as low sulfur fuel and stack gas clean up systems. A 100-200 MW working prototype of the oil gasification process installed in an existing power plant is required to demonstrate the concept.

A three-phase demonstration plant program has been conceived: Phase I - preliminary design, Phase II - detail design and construction and Phase III - developmental operation.

A cooperating utility is needed to carry out such a program. In Phase I the utility would supply information on its power plant, fuel costs, and plant duty cycle. It would participate in the selection of an architect engineer for the preliminary design of a gasifier/desulfurizer system, in the choice of plant options, and in the economic and technical evaluation of the system.

In Phase II the utility would work closely with the architect engineer in the detailed design and installation of the gasifier/desulfurization and share in the costs.

Finally, in Phase III the utility would start-up and operate the plant, collect data on the operation of the gasifier/desulfurizer and cooperate in evaluating the process.

A number of benefits to the utility may be gained from designing, installing, and operating an oil gasifier/desulfurizer system on its existing power plant: reduced emissions of SO_2 and NO_x, elimination of tube deposits and corrosion, capability of burning low grade, low cost resid, and demonstration of leadership in pollution abatement and fuel utilization.

II. OIL FUEL SITUATION

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OIL GASIFICATION/DESULFURIZATION DEVELOPMENT TEAM

Esso (England)
- devised process
- bench scale, batch tests
- pilot operation, continuous tests

Office of Research and Monitoring, EPA
- support
- overall direction

Westinghouse
- systems
- conceptual designs
- evaluations

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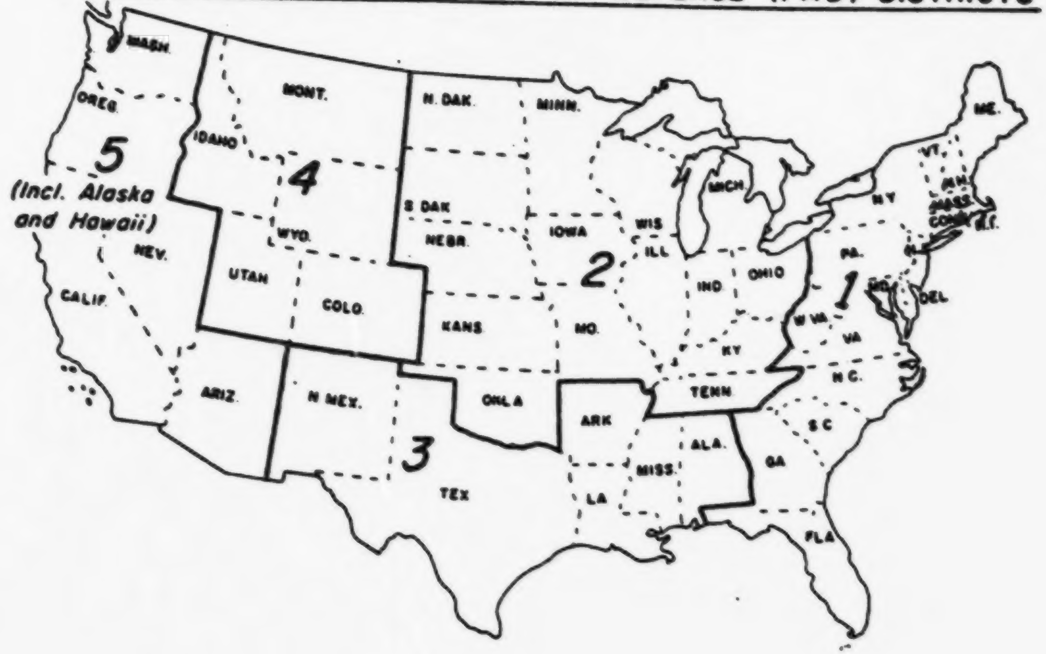
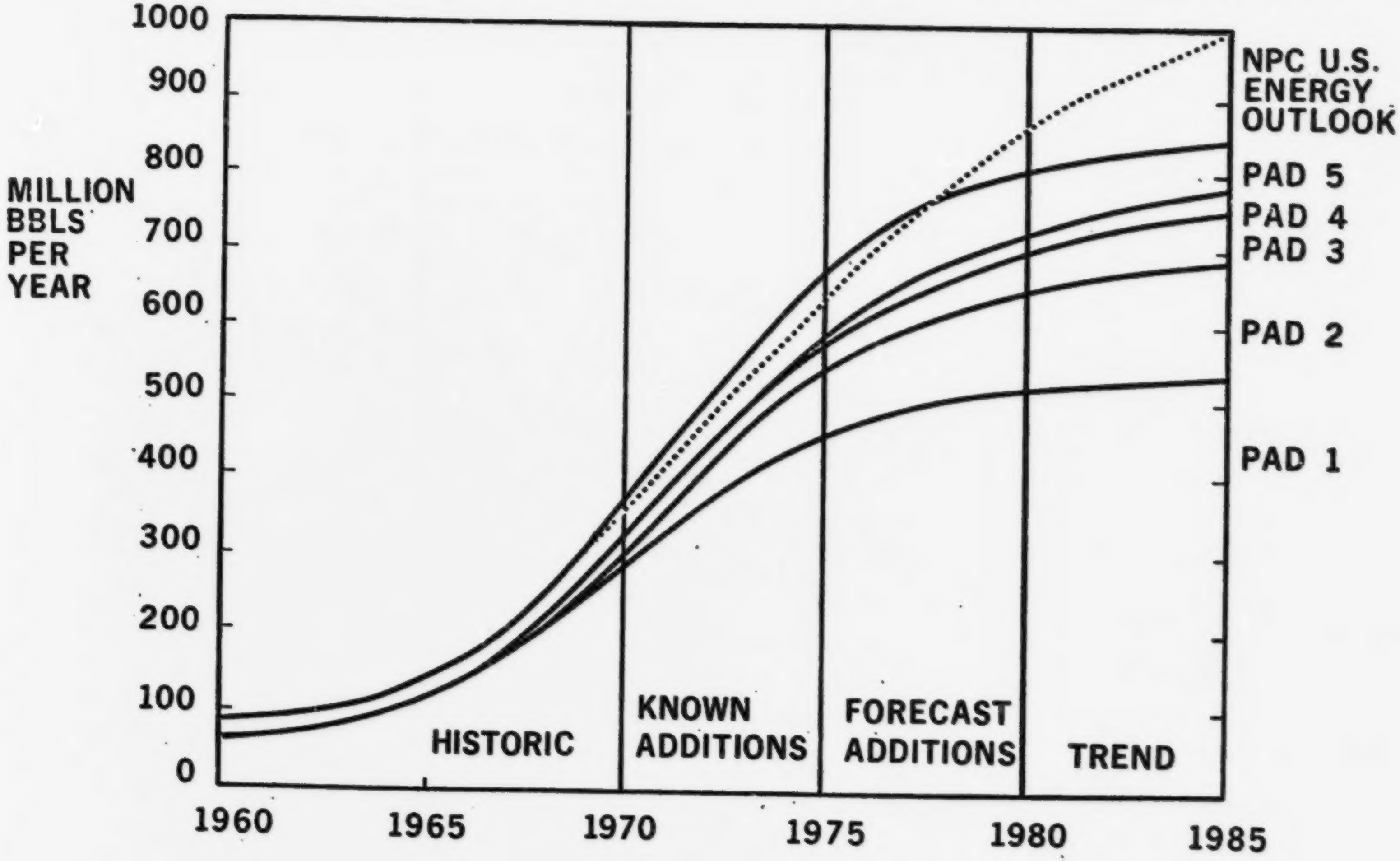
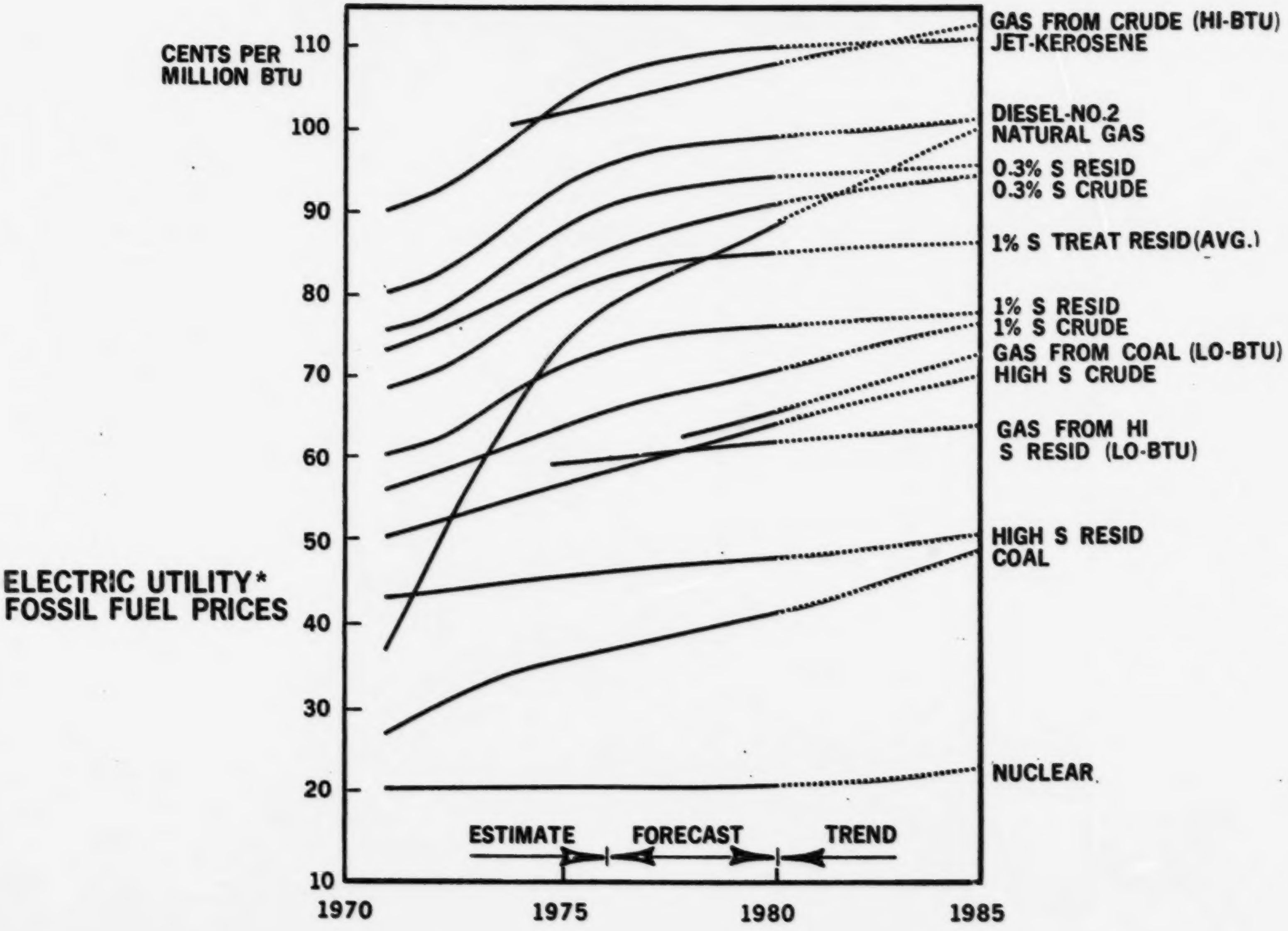
PETROLEUM ADMINISTRATION FOR DEFENSE (PAD) DISTRICTS



ELECTRIC UTILITY FUEL OIL DEMAND BY PAD DISTRICTS*



*January 1972 projection

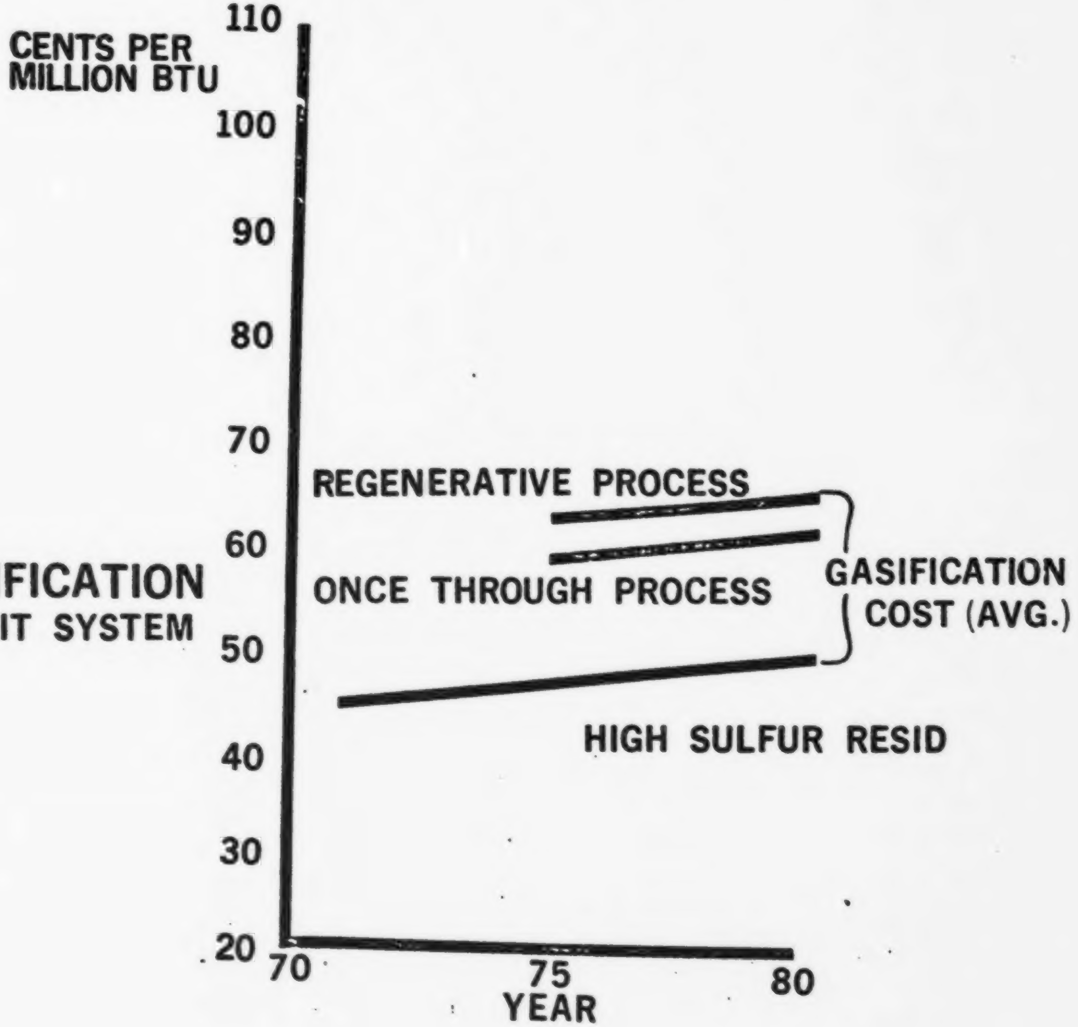


ELECTRIC UTILITY* FOSSIL FUEL PRICES

*January 1972 projection
Does not represent current fuel price situation.

RESIDUAL OIL GASIFICATION

LOW BTU GAS, RETROFIT SYSTEM



III. BASIC PROCESS CONCEPTS

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UTILIZATION OF HIGH-S OIL

In existing plants
- desulfurizing by refinery
- gasifying at power plant
- scrubbing of stack gases

In new plants
- above methods
- burning in fluidized bed boiler
- gasifying under pressure for combined cycle

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GASIFICATION/DESULFURIZATION: CAFB PROCESS

Gasification

$$\text{Oil} + \text{20-25\% Stoichiometric Air} \rightarrow HC + CO + H_2 + CO_2 + H_2O$$

Desulfurization

$$\left\{\begin{matrix} CaO \\ CaCO_3 \end{matrix}\right. + H_2S \rightarrow CaS + H_2O + \left\{\begin{matrix} - \\ CO_2 \end{matrix}\right.$$

Regeneration

$$CaS + \frac{3}{2} O_2 \xrightarrow{1950°F} CaO + SO_2$$

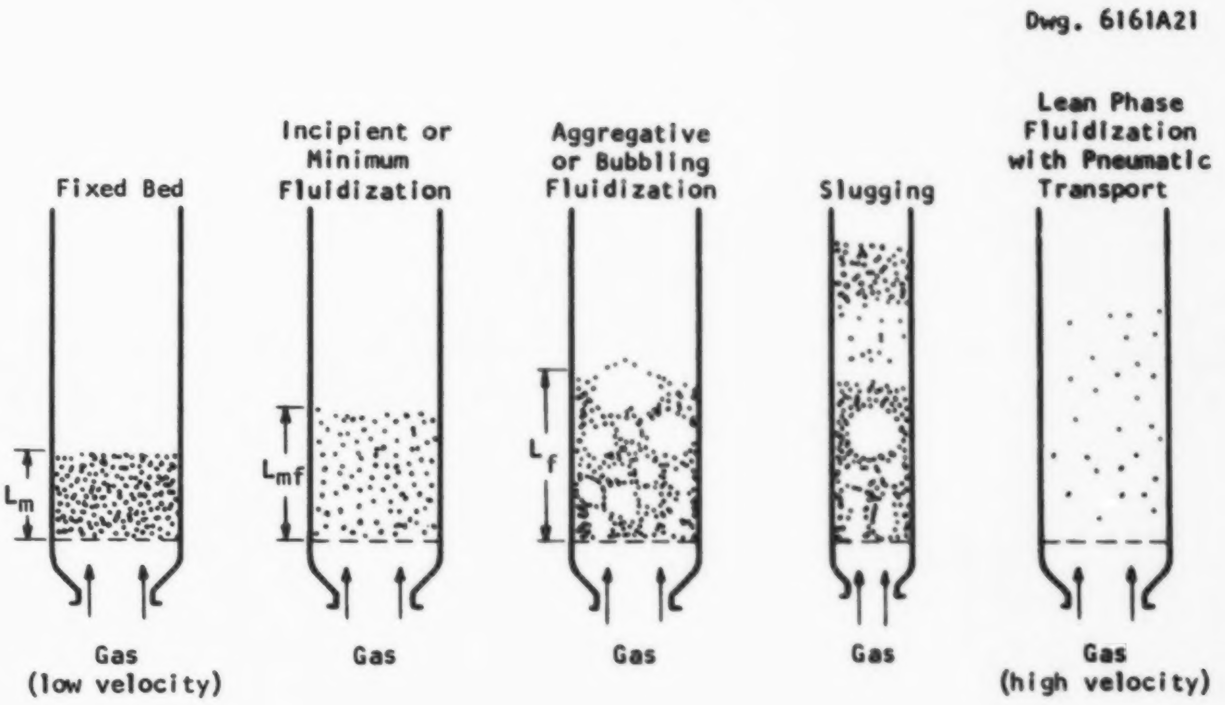
Disposal Alternatives

$$CaS + O_2 \xrightarrow{1700°F} CaSO_4$$

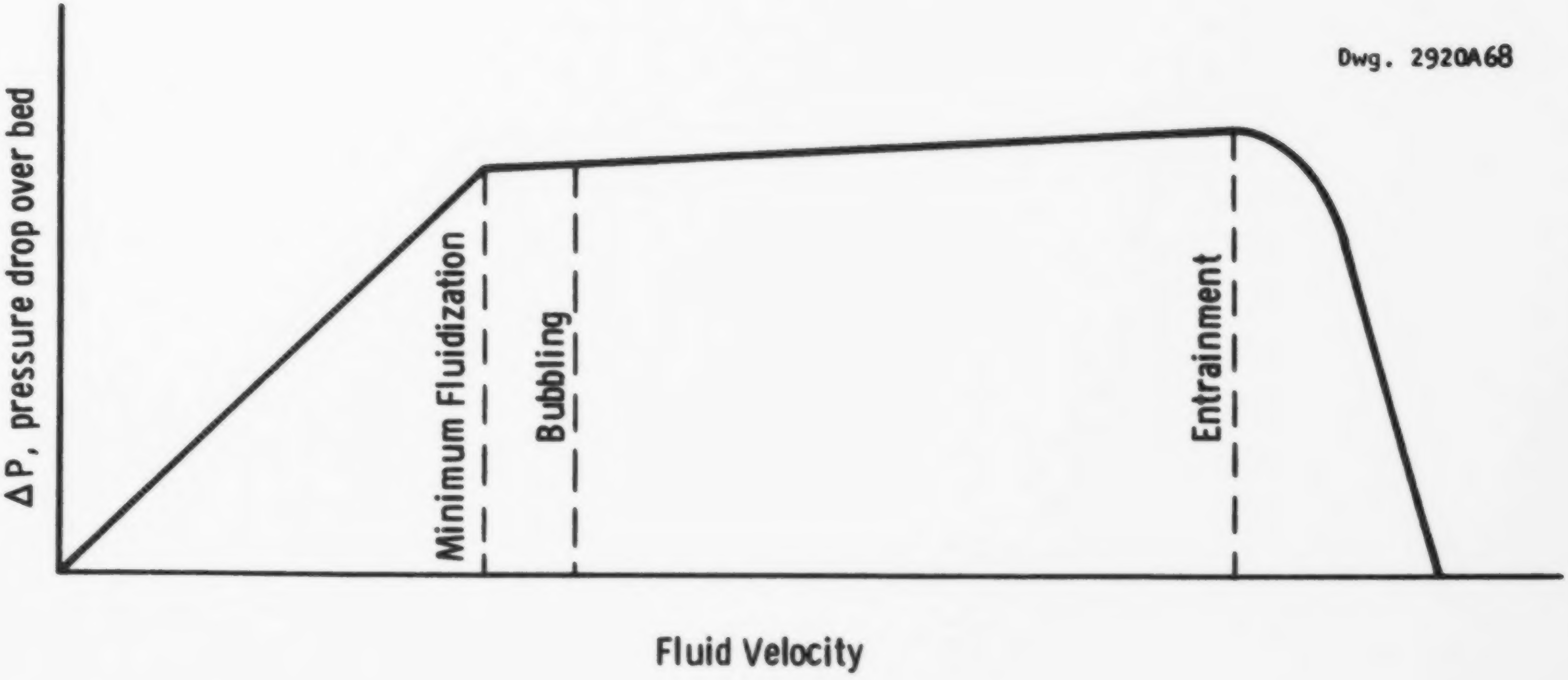
$$CaO + CO_2 + H_2O \rightarrow CaCO_3$$

$$CaO \cdot CaS + H_2SO_4 + H_2O \rightarrow CaSO_4$$

Sintering to unreactive CaO


Dwg. 6161A21
Fixed Bed
Incipient or Minimum Fluidization
Aggregative or Bubbling Fluidization
Slugging
Lean Phase Fluidization with Pneumatic Transport
L_m
L_{mf}
L_f
Gas (low velocity)
Gas
Gas
Gas
Gas (high velocity)

Gas — Solid Contacting



Fluidized bed behavior

CHARACTERISTICS OF FLUIDIZED BEDS

- Liquid-like Behavior - Ease of Solids Handling
- Rapid Solids Mixing - Isothermal Operation
- High Heat Transfer Rates
- Efficient Gas-solid Contacting
- High Gas Through-put

IV. EVALUATION OF FLUIDIZED BED GASIFICATION/DESULFURIZATION

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IV-A. CONCEPTUAL DESIGNS

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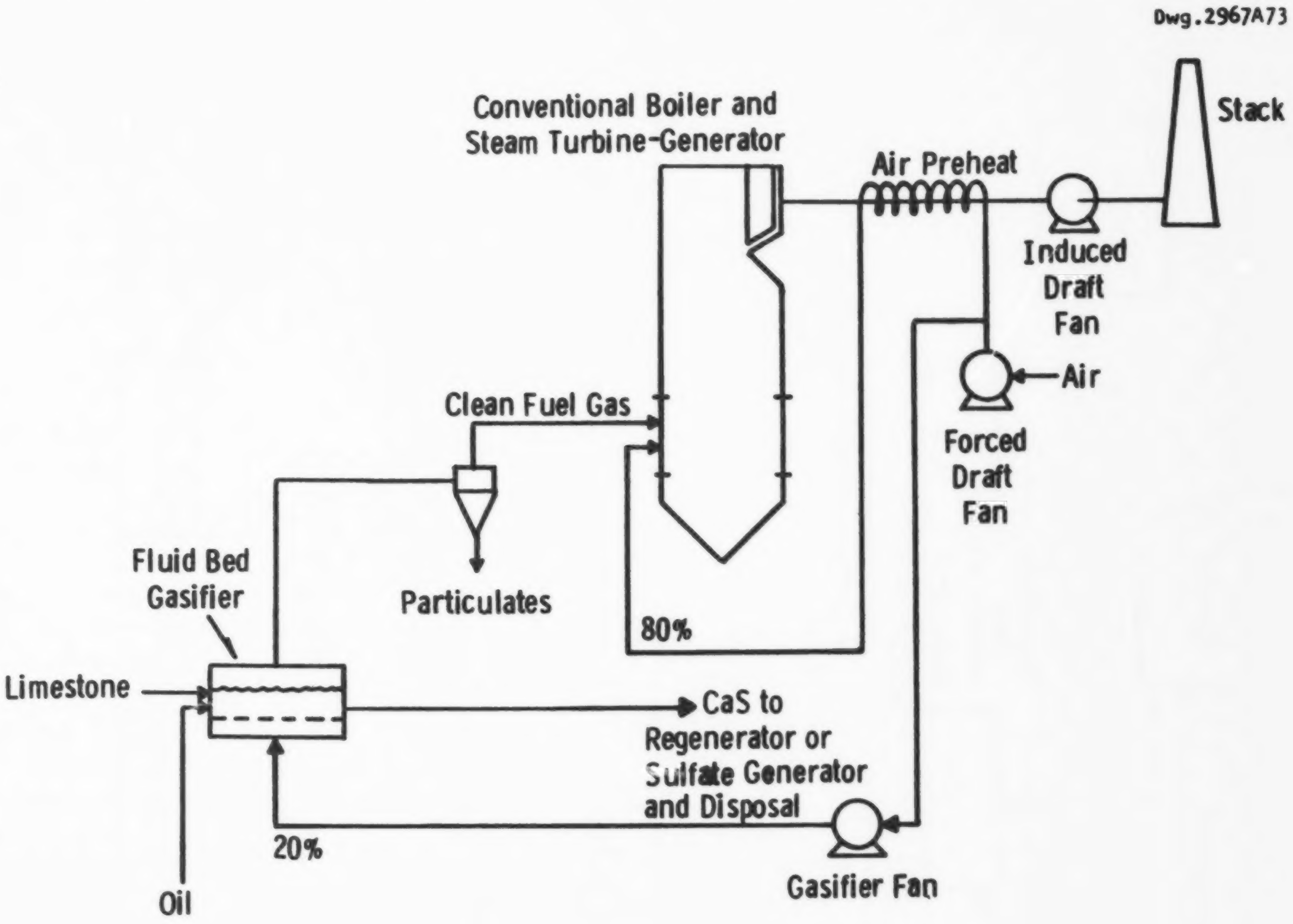

Atmospheric fluidized bed oil gasification power plant

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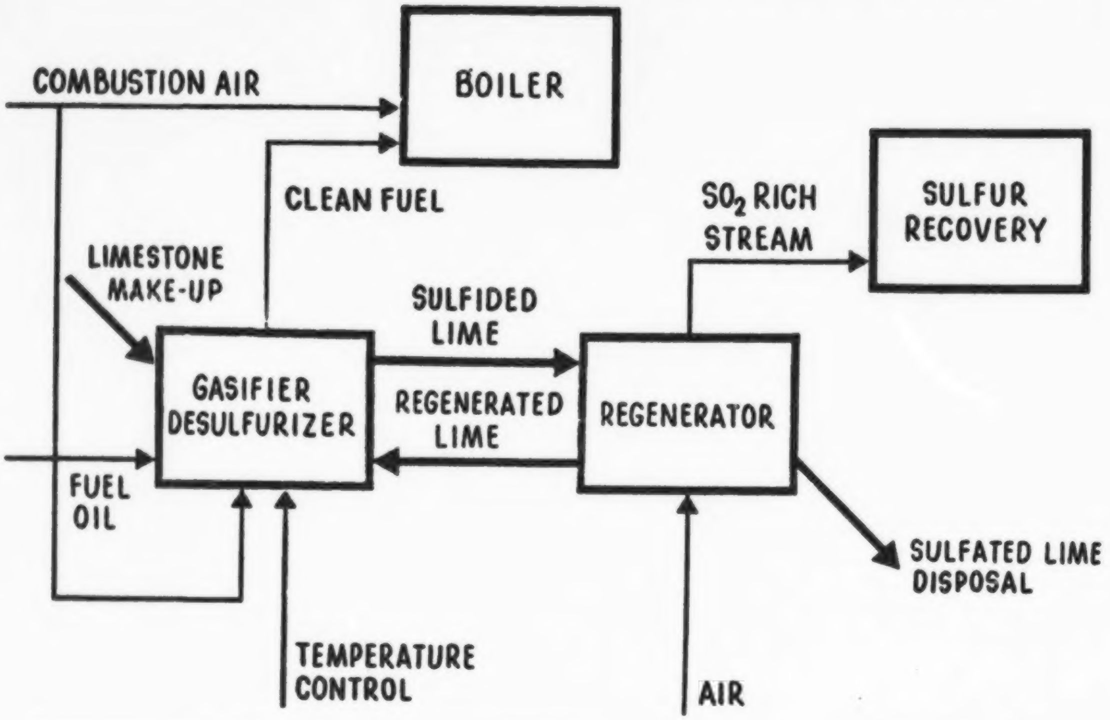
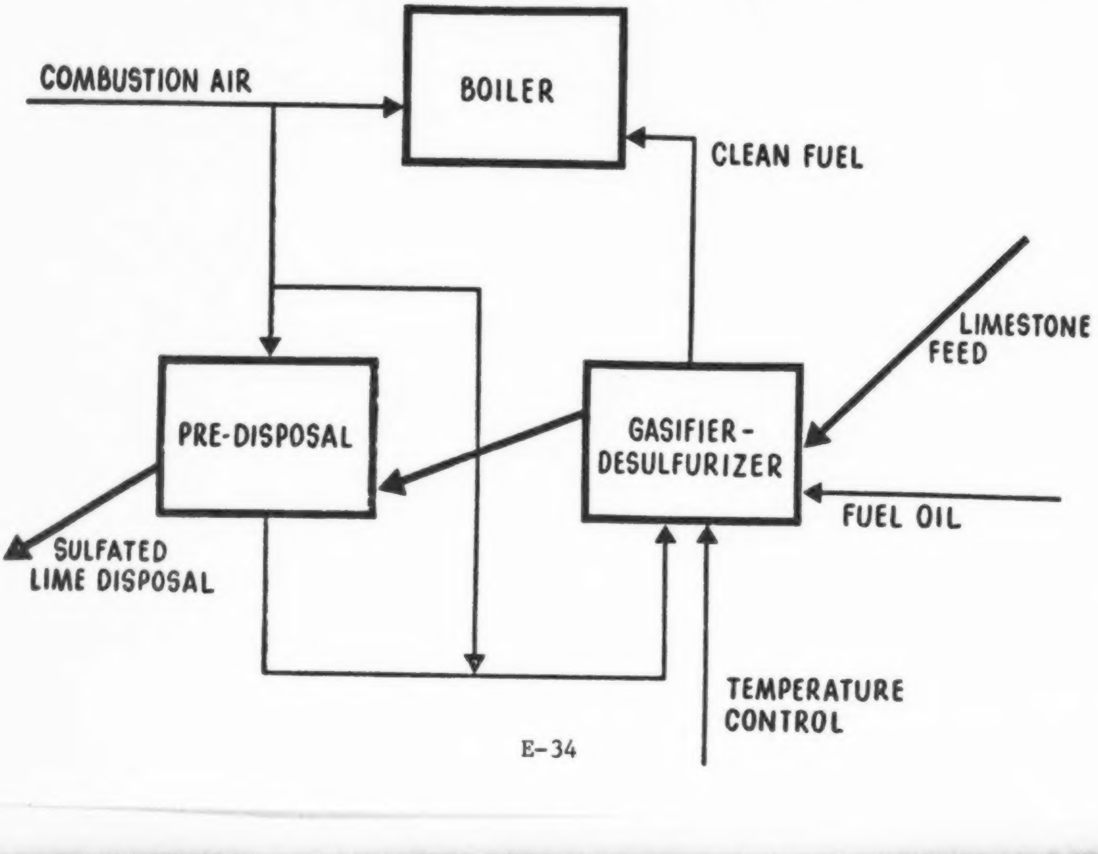
REGENERATIVE OPERATION



ONCE-THROUGH OPERATION



CONCEPTUAL DESIGN

DESIGN BASIS ESTABLISHED

- Pollution Control Requirements
- Functional Specifications
- Operating and Design Parameters
- Economics
- Market

DESIGN CHARACTERISTICS DETERMINED

- Material and Energy Balances
- Equipment Design
- Control
- Plant Layout
- Plant Modification
- Plant Performance

ECONOMICS PROJECTED

- Capital Cost Breakdown
- Fuel Adder Cost

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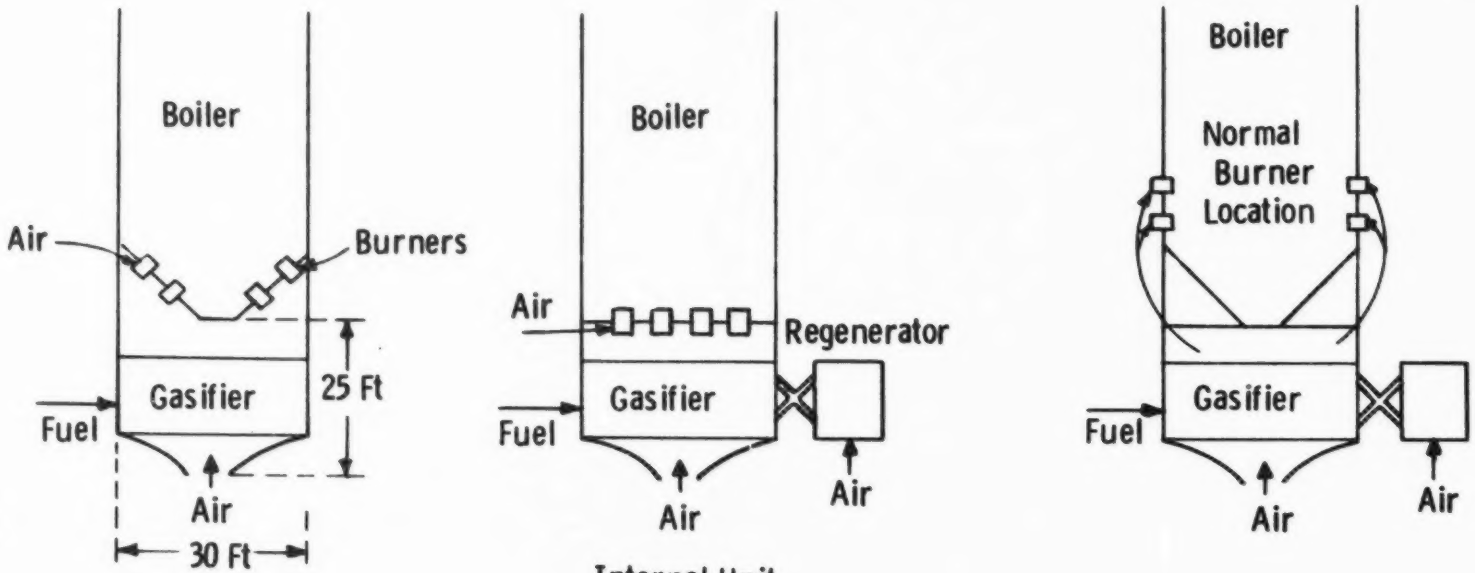
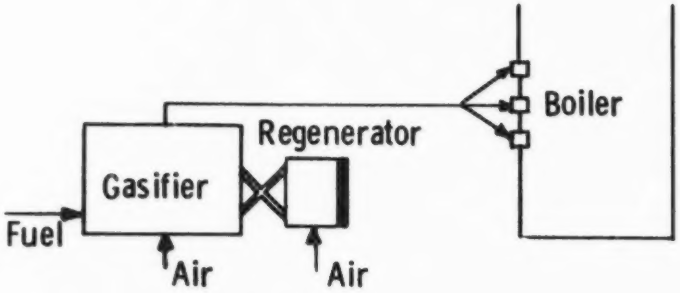


Internal Unit



External Unit

Boiler add-on concepts

FUNCTIONAL DESIGN PARAMETERS

Plant Capacity	600 MW
Location	Eastern U.S.
Fuel Oil	3 wt % S, LHV = 17,700 Btu/lb
Sulfur Removal	Limestone 90-95% Removal Target
Load Factor	40% and 80%
Turn-Down	4:1

OPERATING PARAMETERS

	Regenerative	Once-Through
Gasifier Temperature, °F	1600	1600
Regenerator Temperature, °F	1900	---
Pre-Disposal Unit Temperature, °F	---	1500
Stone Make-up Rate, moles CaO/mole S	1	3
Air/Fuel Ratio, % of stoichiometric	20	20
Mean Limestone Particle Size, μm	2000	1000
Fluidizing Velocity, fps	8	8
Gasifier Bed Depth, ft	2.5-3.5	3.5-4.0

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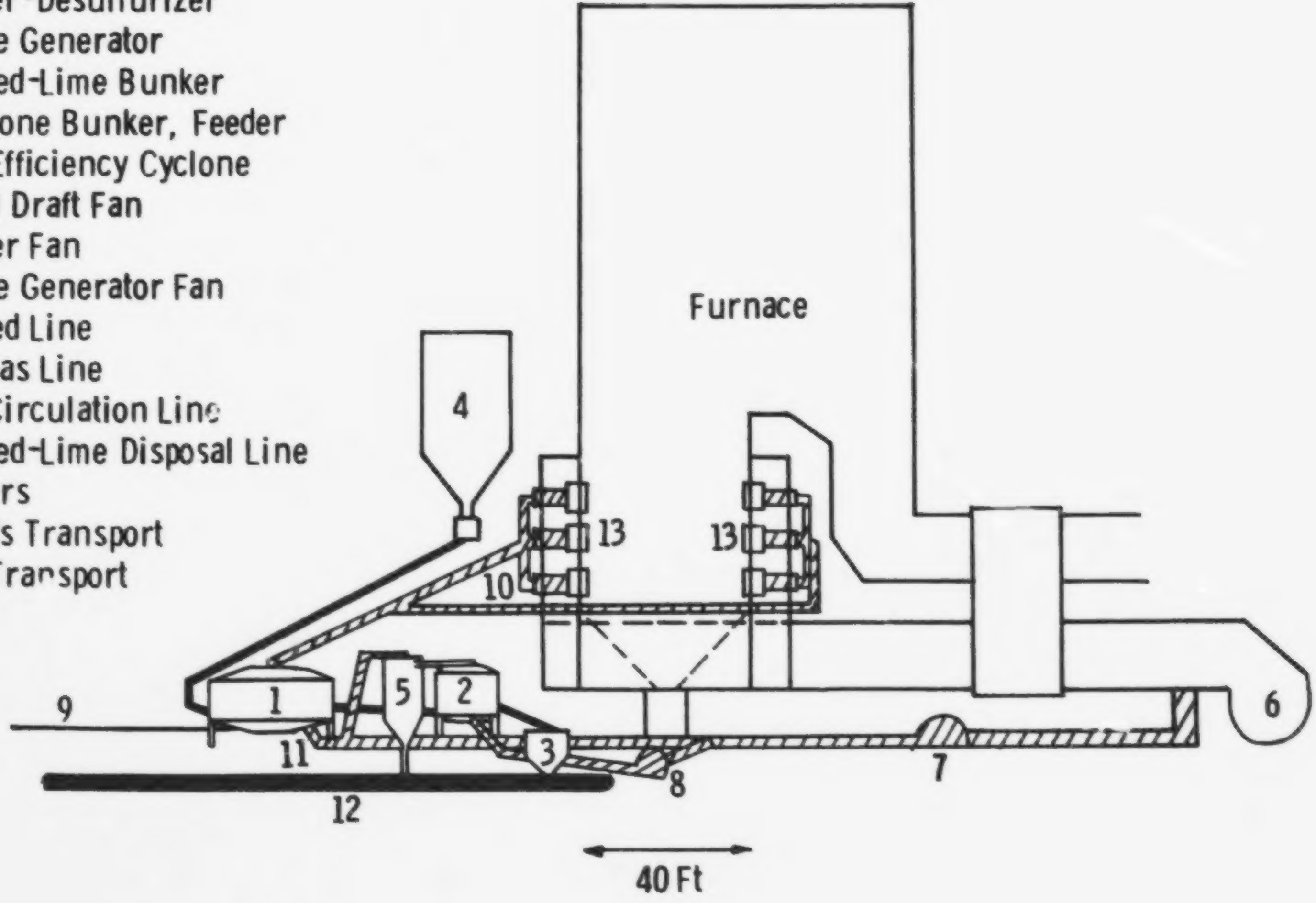
600 MW Boiler Retrofit-
External Once-Through Design

1. Gasifier-Desulfurizer
2. Sulfate Generator
3. Sulfated-Lime Bunker
4. Limestone Bunker, Feeder
5. High-Efficiency Cyclone
6. Forced Draft Fan
7. Gasifier Fan
8. Sulfate Generator Fan
9. Oil Feed Line
10. Fuel Gas Line
11. Lime Circulation Line
12. Sulfated-Lime Disposal Line
13. Burners

▬ Solids Transport
▨ Gas Transport

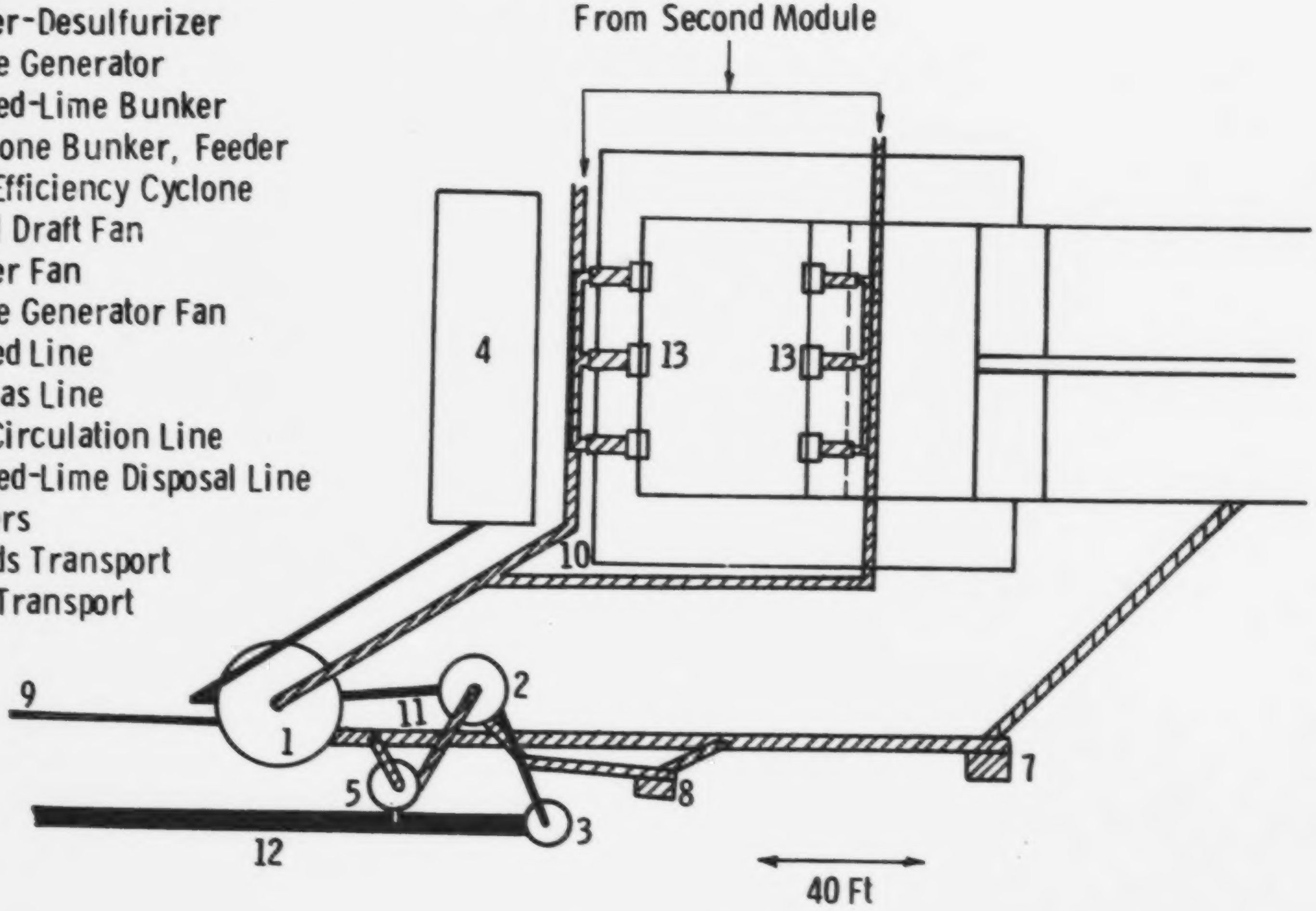


Dwg. 2960A99

600 MW Boiler Retrofit-
External Once-Through Design

1. Gasifier-Desulfurizer
2. Sulfate Generator
3. Sulfated-Lime Bunker
4. Limestone Bunker, Feeder
5. High-Efficiency Cyclone
6. Forced Draft Fan
7. Gasifier Fan
8. Sulfate Generator Fan
9. Oil Feed Line
10. Fuel Gas Line
11. Lime Circulation Line
12. Sulfated-Lime Disposal Line
13. Burners

■ Solids Transport
▨ Gas Transport



IV-B. ASSESSMENT

- PERFORMANCE
- ECONOMICS
- ADVANTAGES
- PROBLEM AREAS

COMPARISON FOR RETROFIT SYSTEM
ONCE-THROUGH VS REGENERATION

	Once-Through	Regeneration
Economics		
Capital cost, \$/kW	25	35
Fuel adder cost, ¢/10^6 Btu	13	16
Space	Gasification system requirements similar Regeneration system requires space for sulfur recovery plant	
Performance		
Efficiency, %	96-97	90-96
SO_2 removal, %	90-95	90-95
Waste disposal ft^3/MW-day	45	15
Operation	Solids circulation problems minimized No sulfur recovery plant	

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CAPITAL COSTS - RETROFIT SYSTEM

. 600 MW Plant
. 3 wt % sulfur oil
Costs in $/kW

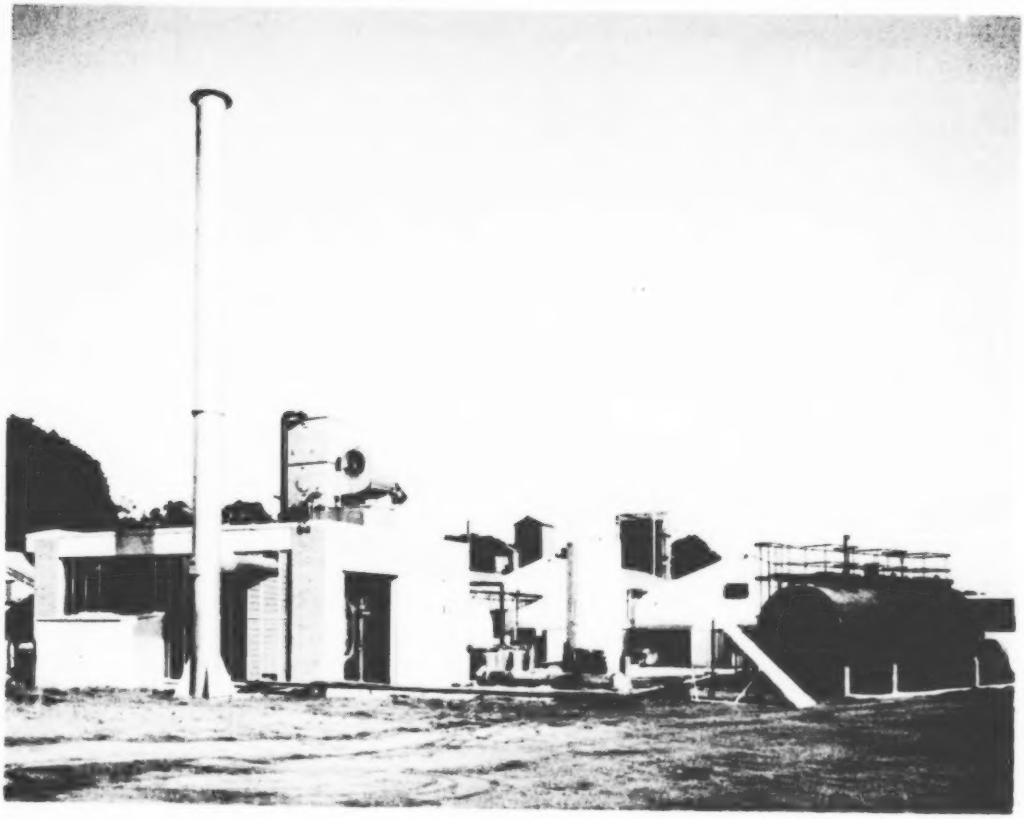
	Oil Gasification		Stack Gas Cleaning	Low Sulfur Oil
	Regenerative	Once-Through		
Cost for:				
SO_2 Control, 0.35 lb/10^6 Btu	16-19	11-13	37-72	0-5
NO_x Minimization, 0.2 lb/10^6 Btu	---	---	2-5	2-5
Particulate Control, 0.1 lb/10^6 Btu	5-7	5-7	---	---
Solid Waste Disposal	1	2	3	---
TOTAL, $/kW	22-27	18-22	42-80	2-10

FUEL ADDER COSTS - RETROFIT SYSTEM

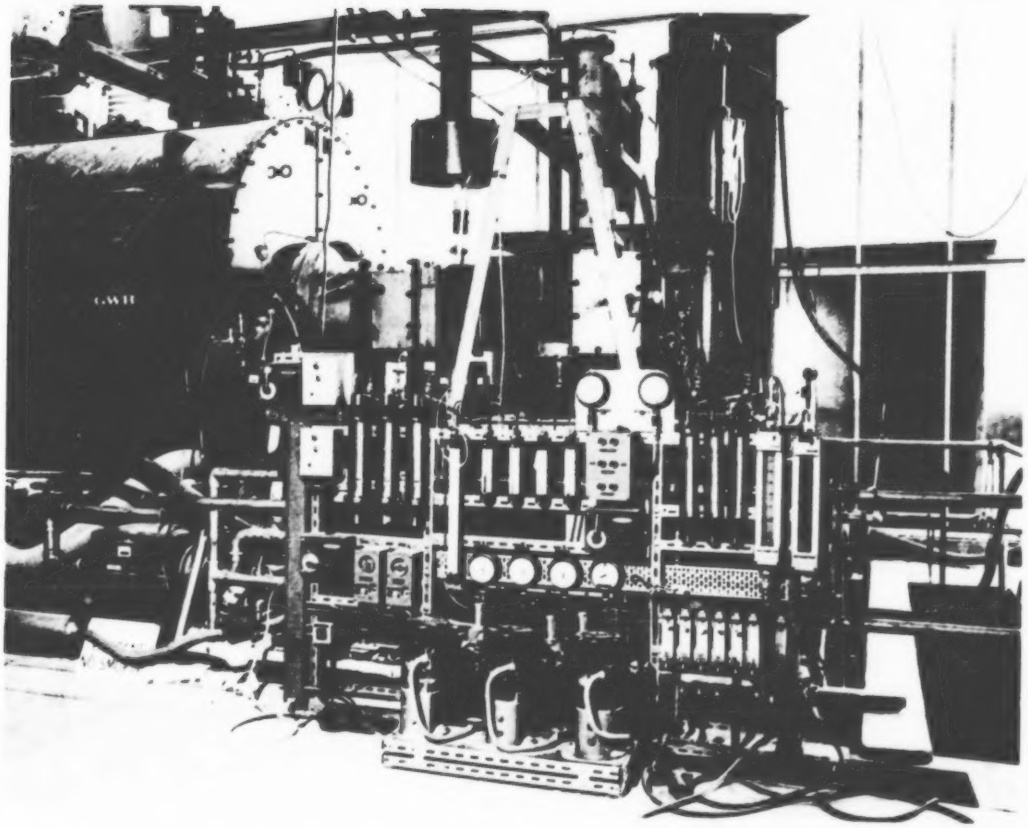
. 600 MW Plant
. 3 wt % Sulfur Oil
. 80% Load Factor
Costs Represented as ¢/10^6 Btu

	Oil Gasification		Stack Gas Cleaning	Low Sulfur Oil
	Regenerative	Once-Through		
Cost for:				
SO_2 Control, 0.35 lb/10^6 Btu	9-15	7-8.5	12-18	17-25
NO_x Minimization, 0.2 lb/10^6 Btu	---	---	0.5-1.5	0.5-1.5
Particulate Control, 0.1 lb/10^6 Btu	1.5-2.5	1.5-2.5	---	---
Solid Waste Disposal	0.5	1.5	2	---
TOTAL, ¢/10^6 Btu	11-18	10-13	14.5-21.5	17.5-26.5

Includes: capitalization, limestone, operating labor, maintenance, efficiency loss, stone disposal, contingency

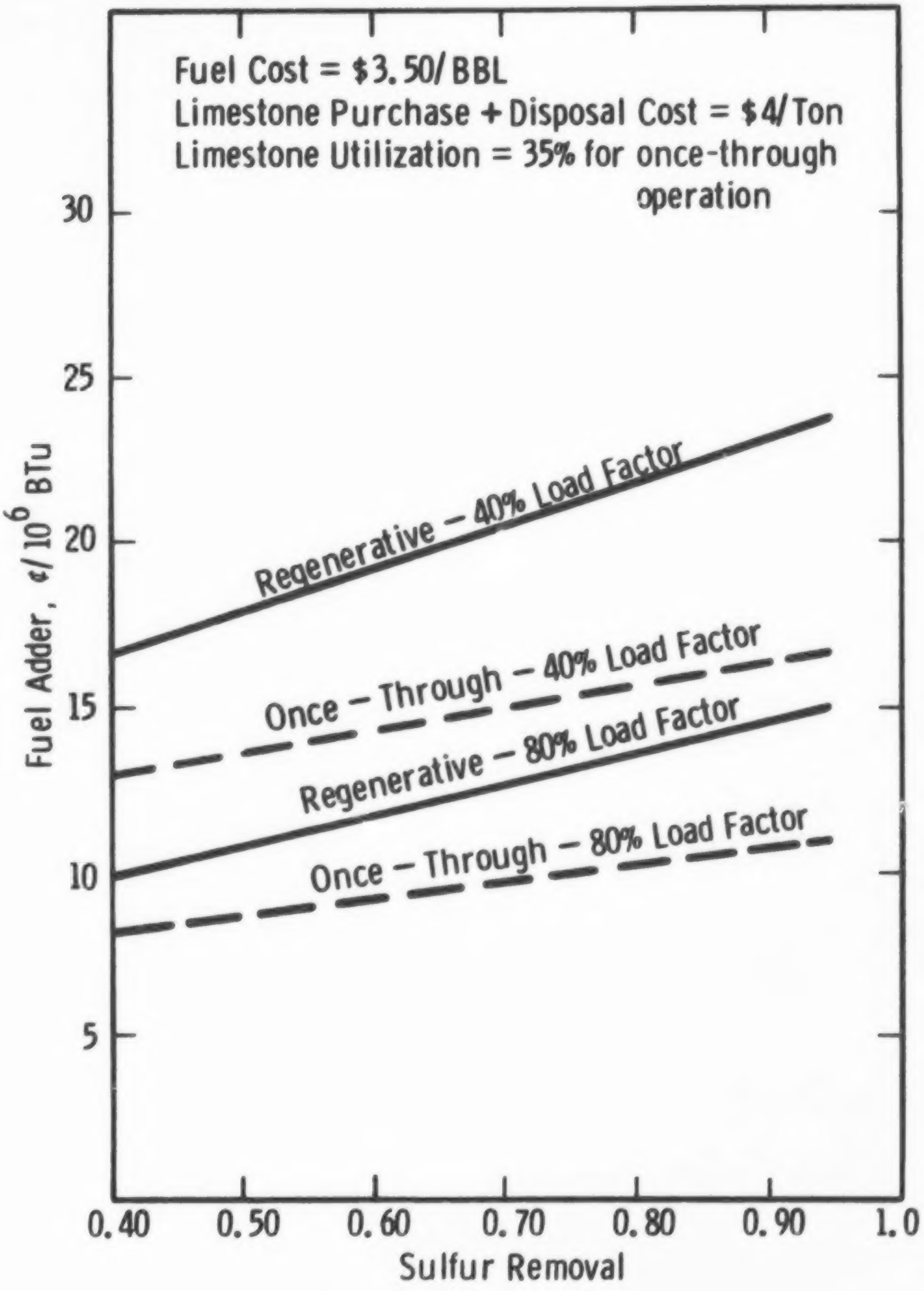


Esso (England) Fluidized Bed Oil Gasification Development Facility



Continuous 750 kW Fluidized Bed Oil Gasifier, Regenerator, and Boiler

Curve 645584-A



Atmospheric-pressure oil gasification fuel adder costs

ADVANTAGES OVER STACK GAS WET SCRUBBERS

- Corrosion and fouling problems minimized in SO_2 removal process and boiler (minimum SO_x and V)
- No flue gas reheat required
- Uses crushed limestone - no limestone pulverizing system needed
- Simplified disposal - potential dry solids and no disposal pond
- More compact system
- Reduced structural costs
- Lower auxiliary power requirement
- Reduced energy cost
- Improved NO_x control
- Potential market for spent CaO

COMPARISON WITH LOW SULFUR FUELS

- Reduced fuel adder
- Increased capital investment
- Availability and stability of low sulfur fuel supply is unreliable
- Improved NO_x control
- Reduction in vanadium and sodium in boiler

PROBLEM AREAS

- Sulfate generator
- Fuel injector number and location
- Turndown
- Sulfur recovery
- Stone circulation

IV-C. EXPERIMENTAL DEVELOPMENT

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DEVELOPMENT PROGRAM

Pilot Scale Experiments Have Demonstrated Concept

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EXPERIMENTAL DEVELOPMENT

ACCOMPLISHMENTS

- Establish sulfur removal capabilities
- Provide design data
- Operate continuous 750 kW gasifier-regenerator system

EXPERIMENTAL APPARATUS

Description	Range of Operating Conditions	Types of Experiments
· 4 batch fluidized beds ·3 with 7" diameter and 5-1/4' height ·1 with 7" diameter and 12-1/2' height ·Other differences were in cyclone location, insulation, etc.	· Temperature = 1500°F-2000°F · Atmospheric pressure · Air/fuel ratios of 15-30% of stoichiometric · Bed depths of 10-20" · Velocities of 3-8 ft/sec · Particle diameters of 500μ-2000μ · Runs of 1-2 hours	· Batch gasification/desulfurization · Batch regeneration
· Continuous unit - cold model ·Constructed of thin steel sheet with transparent plastic solids transfer lines	· Cold unit · Lime circulation rates up to 700 lb/hour	· Lime circulation and control of system
· Continuous unit ·750 kW capacity ·Equipped with boiler ·No sulfur recovery	· 200 hour run at conditions indicated by batch studies	· Continuous and long-term operation demonstration · Future runs will collect specific operating data

EXPERIMENTAL RESULTS

BATCH TESTS

- Nearly 100% sulfur removal at 1600°F, 20% air/fuel ratio, with limestone make-up rate of 1 mole CaO per mole of sulfur fed
- Nearly 100% of fuel vanadium removed
- About 20% of fuel sodium removed
- Clean product fuel is easily combustible with a luminous, smoke free, flame
- Regeneration at 1900°F gives a 10% SO_2 gas product and a regenerated lime having reduced activity

CONTINUOUS UNIT

- Early results indicate superior behavior during continuous operation due to improved temperature control
    - Lower air/fuel ratios, down to 17% of stoichiometric, have been demonstrated
    - 80 to 90% sulfur removal has been demonstrated

IV-D. DEMONSTRATION PLANT

OBJECTIVES OF DEMONSTRATION PLANT STUDIES

- Demonstrate commercial feasibility of process
- Provide detailed design and cost information
- Provide operating experience and performance data

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DEMONSTRATION PLANT

DESIGN CONSIDERATIONS

- Retrofit coal or oil fired plant?
- Capacity?
- Gasifier system location - Integral to boiler or
  External to boiler?
- Regenerative or once-through operation?
- Unit design or modular?
- By-pass requirements?
- Solids disposal
- Temperature control

OPERATION CONSIDERATIONS

- Availability of unit for experimental program?

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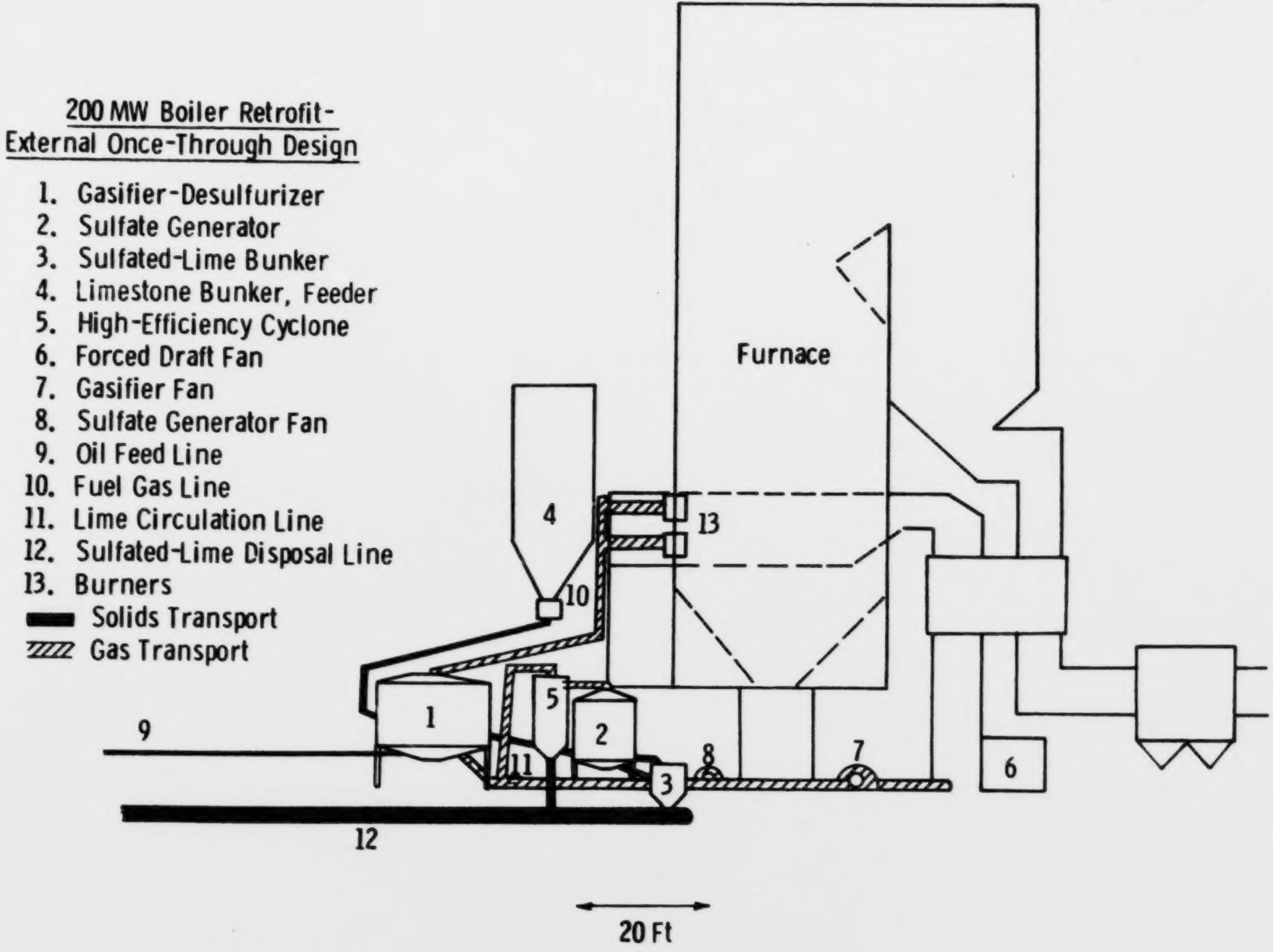

200 MW Boiler Retrofit-
External Once-Through Design

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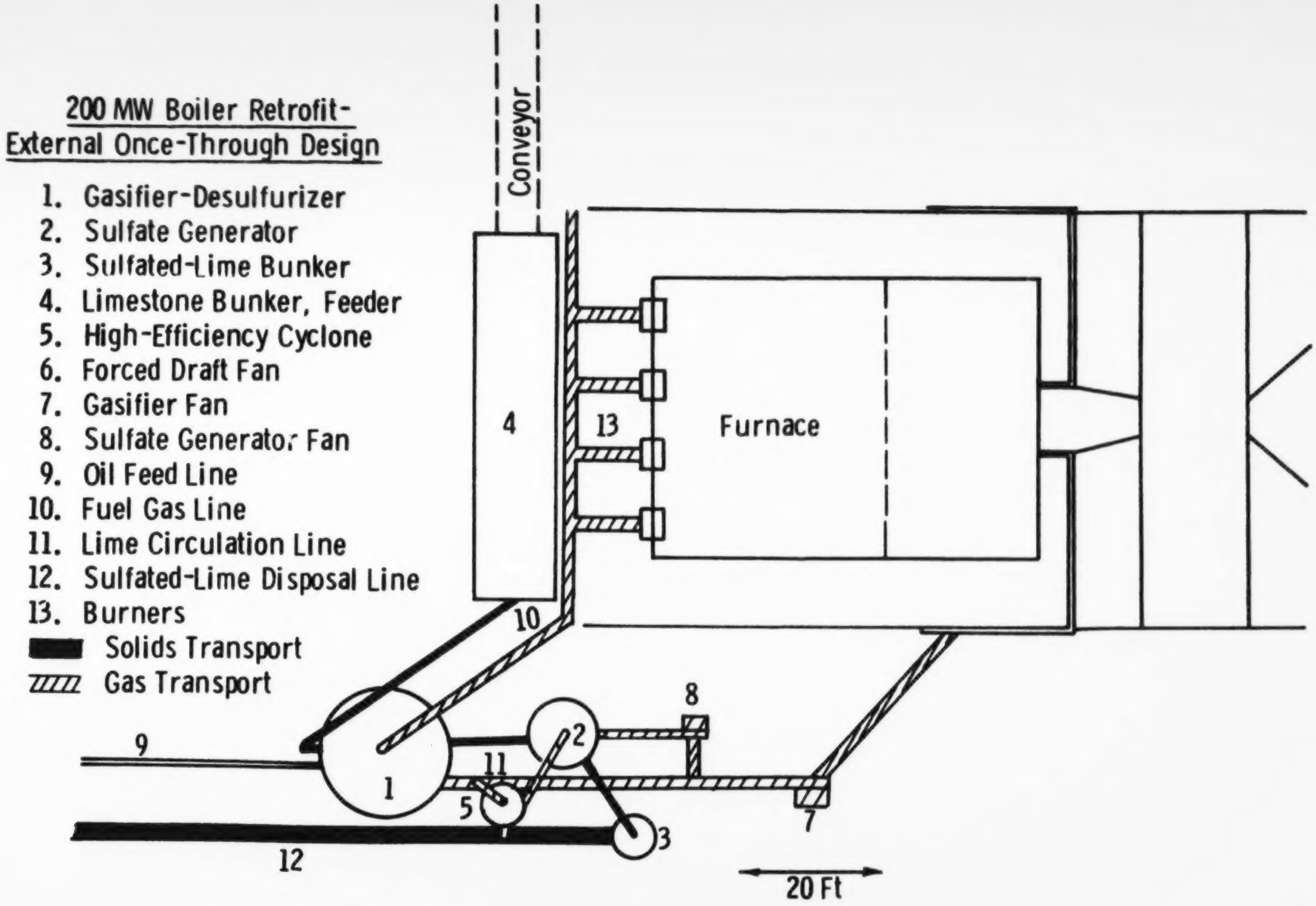
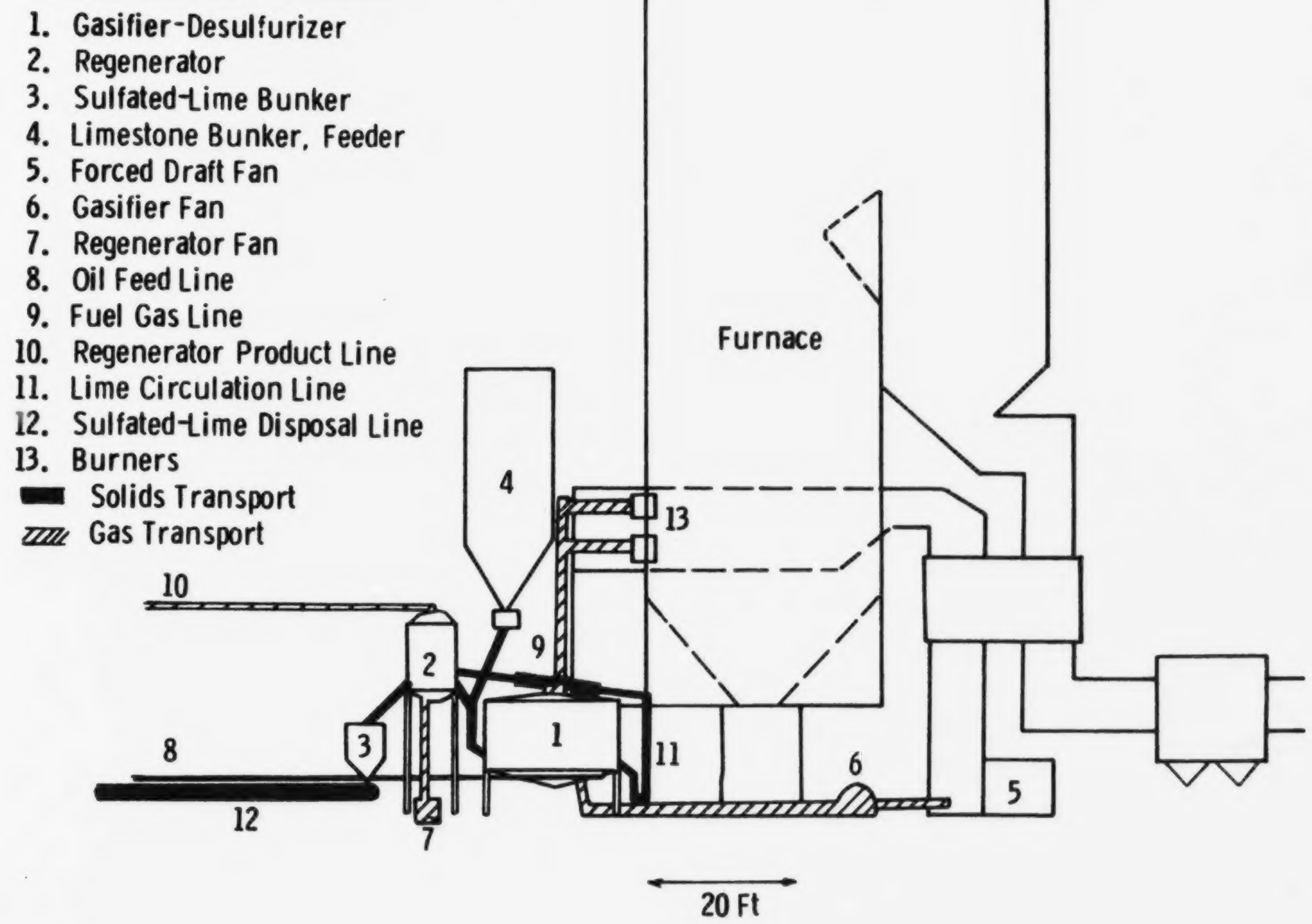

200 MW Boiler Retrofit-
External Once-Through Design
1. Gasifier-Desulfurizer
2. Sulfate Generator
3. Sulfated-Lime Bunker
4. Limestone Bunker, Feeder
5. High-Efficiency Cyclone
6. Forced Draft Fan
7. Gasifier Fan
8. Sulfate Generator Fan
9. Oil Feed Line
10. Fuel Gas Line
11. Lime Circulation Line
12. Sulfated-Lime Disposal Line
13. Burners
▬ Solids Transport
▨ Gas Transport
Conveyor
4
13
Furnace
10
8
9
2
11
1
5
3
7
12
20 Ft

Dwg. 2960A96

200 MW Boiler Retrofit-
External Regenerative Design



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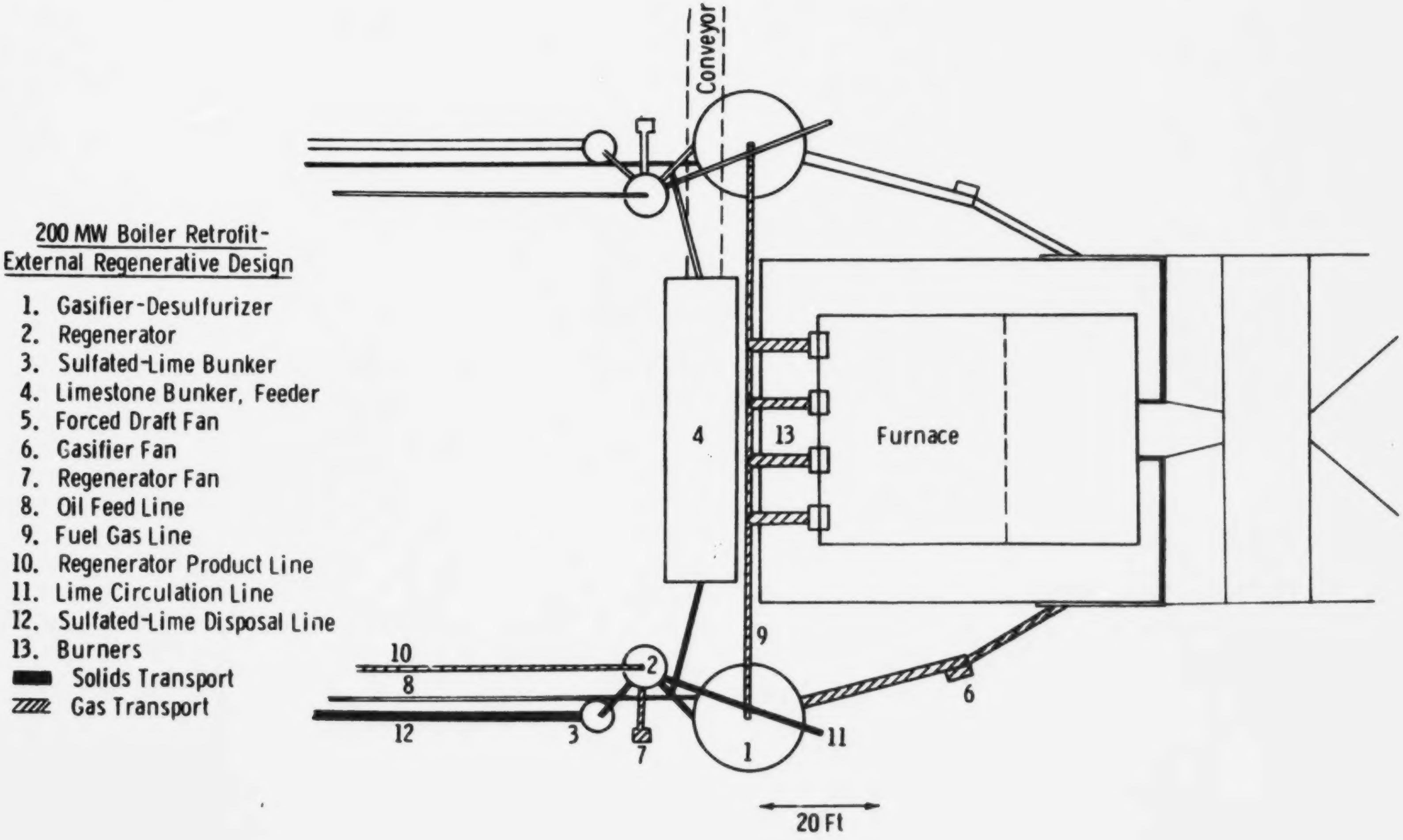

200 MW Boiler Retrofit-
External Regenerative Design

1. Gasifier-Desulfurizer
2. Regenerator
3. Sulfated-Lime Bunker
4. Limestone Bunker, Feeder
5. Forced Draft Fan
6. Gasifier Fan
7. Regenerator Fan
8. Oil Feed Line
9. Fuel Gas Line
10. Regenerator Product Line
11. Lime Circulation Line
12. Sulfated-Lime Disposal Line
13. Burners

■ Solids Transport
▨ Gas Transport

V. PROPOSED DEMONSTRATION PLANT PROGRAM

PROPOSED GASIFICATION/DESULFURIZATION DEVELOPMENT-DEMONSTRATION PLANT PROGRAM

Utility	Esso	AE	ORM
Phase I - preliminary design, cost and performance estimate			
Information plant fuels generation Selection AE plant options Evaluation	Information (Esso) process system Supervision AE Evaluation	Preliminary plant Cost estimates detailed design equipment installation Performance evaluation	Support Guidance Evaluation
Decision on program continuation: proposal to OAP			
Phase II - detail designs, plant installation			
Supervision Plant site preparation Support	Developmental operation planning Consultation Checking	Plans Procurement Installation	Support Guidance
Phase III - plant start-up, operation, data gathering			
Start-up Plant operation Data gathering Evaluation	Assistance in start-up and operation Data analysis Evaluation	Plant modification	Support Guidance Evaluation
Report on performance and evaluation			

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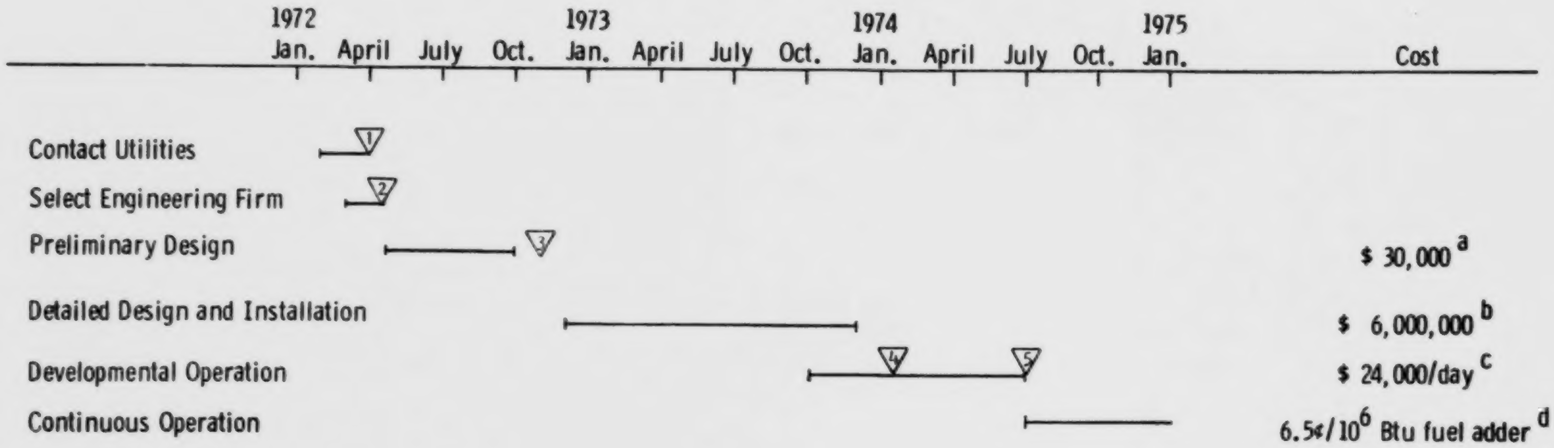
Dwg. 7238675

DEMONSTRATION PLANT SCHEDULE*

Atmospheric Oil Gasification/Desulfurization Process for Existing Steam Power Plants



Milestones

- ▽1 Cooperating utility identified
- ▽2 Engineering firm contract signed for preliminary design
- ▽3 Decision on construction; agreement between OAP - utility - Ⓦ - AE
- ▽4 Shake down complete
- ▽5 Engineering and economic data for future operations complete

Footnotes

a Cost estimate for preliminary design
b Assumes complete 100 MW gasifier module, once-through limestone system
c Maximum cost per day assumes continuous oil feed, O & M for gasifier development, and no power generation
d Based on conceptual design

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*Schedule as projected January 1972. Preliminary design engineering actually initiated September 1973.

APPENDIX F

UTILITY PROSPECTS FOR DEMONSTRATION PLANT

OIL GASIFICATION PROSPECTS

	SCE	LADWP	PG&E	MSU	FP&L	JEA	BG&E	PE
1. High S residual oil available, cheap or exported (Cal. 30,000 bbl/d, Tex 5000 bbl/d)	X	X	X	X	X	X	X	X
2. Have capability of receiving high S residual (ship or PL)	X	X	X	X	X	X	X	X
3. Have serious environmental problems	X	X	X		X	X	X	X
4. Management aggressive in anti-pollution activities	X	X	X	X	X	X	X	X
5. Interested in SO_2 removal systems, crude, etc.	X	X		X	X	X	X	X
6. Interested in Ford, Bacon, Davis crude topping and naphtha							X	
7. Currently using high S residual but must go to low S			X	X	X			X
8. Gas boilers on oil present serious fouling problem				X				
9. Have been cooperative with Ⓦ on oil and SO_2 studies	X	X	X	X	X	X		X
10. Can get money to do R&D work		X				X		

Other candidates -- Boston Edison, PSE&L, JC, FPC, TU, SDG&E

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APPENDIX G

UTILITIES VISITED

	Initial Presentation
Baltimore Gas and Electric	3/7/72
Jacksonville Electric Authority	3/16/72
Florida Power and Light	3/17/72
Florida Power Corporation	3/17/72
Middle South Utilities	4/6/72
Pacific Gas and Electric	5/4/72
Los Angeles Department of Water and Power	5/4/72
Southern California Edison	5/5/72
Savannah Electric and Power	5/18/72
Boston Edison	7/11/72
New England Gas and Electric Association	7/11/72
New England Electric System	7/11/72
Montaup Electric Co.	7/11/72
Commonwealth Edison	7/19/72

APPENDIX H

PROPOSED UTILITY AGREEMENT

AGREEMENT AMONG A UTILITY, WESTINGHOUSE, AND THE OFFICE OF AIR PROGRAMS* (OAP) FOR THE DEVELOPMENT OF ATMOSPHERIC PRESSURE FLUIDIZED BED OIL GASIFICATION

The Office of Air Programs (OAP) of the Environmental Protection Agency (EPA) has undertaken a program to develop an atmospheric-pressure fluidized bed oil gasification/desulfurization process to provide clean fuel to a boiler. Under contract to OAP, Esso Petroleum Co. (England) has collected basic process data. Its effort has culminated in the construction and operation of one megawatt pilot gasifier. Utilizing Esso data, Westinghouse has produced and evaluated conceptual designs of oil gasification/desulfurization systems for full-scale power plants. The concept appears technically feasible; effective in reducing sulfur dioxide, NO_x, and particulate emissions; and economically competitive with alternative sulfur control options. A 100 to 200 MW working prototype of the oil gasification process installed in an existing power plant is required to demonstrate the concept.

A demonstration plant program has been conceived; it is divided into three phases:

- Phase I - Preliminary design and cost estimate of demonstration plant installed on an existing boiler
- Phase II - Detail design and construction of demonstration oil gasification process
- Phase III - Developmental operation of the gasification process and integrated power plant.

A utility is required as a cooperating party to carry out this program. In Phase I the utility will

- Supply technical information on its existing power plant -- plant layouts, equipment drawings, design and operating data

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*The Office of Air Programs became the Office of Research and Monitoring and is now the Office of Research and Development.

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- provide price and availability information on oil fuels which might be utilized in the plant currently and/or in the future.
- give information concerning the generation requirements placed on the plant -- daily and seasonal variations in loading; availability of the plant for modification and developmental tests.
- cooperate in the selection of an engineering firm to prepare preliminary drawings, cost estimates and schedules for the installation of an oil gasifier/desulfurizer.
- assist in selecting various options in the design of the system -- location of the equipment, choice of burners, means of bypass, technique for temperature control, etc.
- evaluate in cooperation with Westinghouse and OAP the effectiveness and economy of the oil gasifier/desulfurizer in power generation and pollution abatement.

Westinghouse, under contract to OAP, will direct the work of the engineering firm as it prepares, under subcontract to Westinghouse, the preliminary gasifier/desulfurizer system designs and cost estimates for the utility power plant. OAP will provide general guidance and funding for Phase I.

The engineering firm will carry out designs in sufficient detail that fixed price bids can be solicited for detailed design and construction of the system. Specifically the information developed during Phase I will include

- evaluation and selection of process options (e.g., once-through vs. regenerative use of limestone method of temperature control)
- process and engineering flow diagram
- equipment list
- material and equipment specifications
- plot plans and layout drawings
- perspective drawings

- the utilities requirement for operation, availability, etc.
- operating program for the development phase
- cost estimates for detail design and construction and for developmental operation
- schedule

The design effort will be based on experimental data from Esso (England) on their one MW continuous pilot plant and the conceptual design and assessment by Westinghouse.

At the conclusion of Phase I all participants in the program -- the utility, Westinghouse, and OAP -- will mutually consider whether the program should be carried further. If the information developed in Phase I confirms the effectiveness of the system, the utility and Westinghouse will jointly prepare a proposal to OAP for extension of the work to Phases II and III and will seek funds from industry sources to supplement those available from OAP.

In Phase II the utility will

- work closely with the engineering firm in the design and installation of the gasifier/desulfurizer system.
- share in the costs of the design and installation.

Westinghouse will assist in providing technical data and in checking plans. OAP will continue to provide general guidance and will financially support up to 33% of the project costs.

Finally in Phase III the utility will

- start up and operate the plant
- collect basic data on the operation of the gasifier/desulfurizer system and aid in its interpretation and analysis.
- cooperate with Westinghouse and OAP in the evaluation of the effectiveness, technical and economic, of the process in power generation and pollution abatement.

Westinghouse will provide a suggested developmental test program to obtain data from the system necessary for optimizing its operation, for evaluating its performance, and for designing future plants. OAP will continue to provide partial support for Phase III.

At the conclusion of the program a joint report will be prepared by the utility and Westinghouse summarizing system designs, operating data, and performance evaluations. Both the value of the system to the utility and to the industry will be considered. The results of the program will be presented to the utility, Westinghouse, and OAP management, and to the industry.

A number of benefits to the utility may be gained from designing, installing, and operating an oil gasifier/desulfurizer system on an existing power plant.

- greatly reduced emissions of SO_2 and NO_x in burning high sulfur oil
- elimination of tube deposition or corrosion in the burning of high vanadium resids
- capability of burning low grade, low cost petroleum residual stocks with a minimum of combustion difficulties
- demonstration of utility leadership in pollution abatement and effective fuel utilization
- sharing of financial and technical risks in installing and operating a system based on new technology.

In view of these benefits the utility agrees to participate in the program as described above and summarized in the attached schedule. The utility will provide technical data and engineering manpower to complete its tasks and evaluations of Phase I. If technical and economic prospects of the installation appear favorable to utility management, it will cooperate in making a joint proposal to OAP for completion of Phase II and III of the program as described.

APPENDIX I

UTILITY MANPOWER REQUIREMENT FOR PHASE I

Utility Participation
in the
DEMONSTRATION PLANT PROGRAM
FOR
ATMOSPHERIC PRESSURE FLUIDIZED BED OIL GASIFICATION/DESULFURIZATION

A 50 to 200 MW working prototype of the oil gasification process installed in an existing power plant is being proposed.(a) A utility is required as a cooperating party to carry out this program. A three-phase demonstration plant program is proposed with the Office of Air Programs (OAP), a utility, an architect engineer, Westinghouse, and with supporting data from Esso (England). The division of responsibilities for the three-phase program is summarized in the attached matrix.

The Phase I preliminary design, cost and performance estimate provide the basis for a critical evaluation of the process and a proposal to OAP for extension of the work. Westinghouse, under contract to OAP, will direct the work of the engineering firm as it prepares the preliminary gasifier/desulfurizer system designs and cost estimates. Phase I design, evaluation, and the formulation of a proposal are estimated to require 8 to 12 months. Utility participation and estimated manpower requirements during this phase include

	Estimated Man-months of Utility Effort(b)
Assist in selection of AE	0.5
Assist in selection of various process designs and operating options	2.0
Supply technical information - plant layouts, design and operating data, fuel data, generation requirements	2.5
Assist with evaluation	2.0
Assist with proposal to OAP for extension of work (if evaluation confirms effectiveness)	2.0
	9.0

Financial support for the preliminary design drawings will be provided by OAP.

In Phase II the utility would work closely with the architect engineer in the detailed design and installation of the gasifier/desulfurizer system. Phase III would include start-up and operation of the plant. Financial support for phases II and III would be shared between OAP and the utility. The architect engineer and equipment vendors may provide services and special equipment at a reduced cost for the demonstration plant.

(a) Technical evaluation and a suggested agreement among a utility, Westinghouse and the Office of Air Programs are presented in separate documents.

(b) The utility may wish to provide additional man-power and increased participation.

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APPENDIX J

FOLLOW-UP MEETING AGENDA

CLEAN BOILER FUEL OIL GASIFICATION/DESULFURIZATION

Follow-up Meeting at Utilities
after Initial Presentation

INTRODUCTION

Purpose of visit

- Not to "sell" process
- Not to rush consideration
- To determine current status
- To understand current attitude about process and proposition

REVIEW

- Process (refer to blue book)
  - Oil gasification/desulfurization in lime bed
  - Development by Esso (UK), Ⓦ, EPA
  - Provision of hot, clean fuel gas to existing boiler (refer to boiler diagram)
  - Enabling use of high-S, high-V resids with moderate cost added (refer economic chart)
  - Sulfur recovery/disposal
- Program (refer to agreement, chart, and schedule)
  - Three-phases: study, construction, operation
  - Costs shared

UTILITY ATTITUDE FROM PRESENTATION AND CONSIDERATION

- Impression of process
  - Technical aspects
    - Workable ?
    - Scalable ?
    - Controllable ?

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- Economic advantages
  - Realistic ?
  - Over-optimistic ?: load factor; equipment cost; installation cost; lime costs; boiler derate; efficiency
- Applicability
  - Availability of oil and lime
  - Convertibility of existing plants
  - Alternate sulfur control measures: once-through regeneration S/H_2SO_4
  - Availability of disposal means
- Impression of proposition
  - Study phase
  - Installation and operation phases

FURTHER STUDY AFTER PRESENTATION

- Who carried this out
- How much effort
- On what did the study focus
- What conclusions

CURRENT STATUS

- Under active consideration
- Awaiting further developments
- Shelved
- Dead

OIL USAGE

- Current
  - Supply source
  - Characteristics S, V
  - Price
  - Terms of commitment
  - Type generation -- base, intermediate, peak

- Future
    - All above
    - Expansion in oil-fired capacity and generation

SO_2 AND NO EMISSIONS

- Current problem in meeting
- Approaches adapted
    - Tactical
    - Long-term approaches

AVAILABILITY OF RESOURCES FOR PROGRAM

- Technical manpower
    - Study
    - Installation and operation
- Research and development funds
- Suitable power plant site for demonstration installation
    - Space
    - Availability for conversion

WILLINGNESS TO COOPERATE IN FURTHERING PROGRAM

- Carrying out demonstration plant installation (\$2.4 million)
- Contributor along with 5 to 6 other utilities installation elsewhere (\$500-1,000 thousand)
- Sponsor of proposal to EPA-ERC for installation (\$25 thousand)
- Sponsor of proposal to EPA-ERC for the development of a pressurized oil gasifier to operate with low-cost combined cycle plant (\$25 thousand)

APPENDIX K

LETTER OF INTENT

NEW ENGLAND ELECTRIC SYSTEM
20 TURNPIKE ROAD
WESTBOROUGH, MASSACHUSETTS 01581
TELEPHONE 617-366-9011

December 12, 1972

Environmental Protection Agency
Control Systems Division
Research Triangle Park
North Carolina 27711

Attention - Dr. John K. Burchard

Gentlemen:

The Office of Research & Monitoring, Control Systems Division of the Environmental Protection Agency, EPA, has undertaken a program to develop an atmospheric pressure fluidized bed oil gasification/desulfurization process to provide clean fuel to a boiler. Under contract to EPA, Esso Petroleum Co. (England) has collected basic process data. Its effort has culminated in the construction and operation of one megawatt pilot gasifier. Utilizing Esso data Westinghouse has produced and evaluated conceptual designs of oil gasification/desulfurization systems for full scale power plants. The concept appears technically feasible; effective in reducing sulfur dioxide, NO_x, and particulate emissions; and economically competitive with alternative sulfur control options. A working prototype of the oil gasification process installed in an existing power plant is required to demonstrate the concept.

A demonstration plant program has been conceived, and it is divided into three phases:

PHASE I - Preliminary design and cost estimate of demonstration plant installed on an existing boiler

PHASE II - Detail design and construction of demonstration oil gasification process

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THE NAME "NEW ENGLAND ELECTRIC SYSTEM" MEANS THE TRUSTEE OR TRUSTEES FOR THE TIME BEING (AS TRUSTEE OR TRUSTEES BUT NOT PERSONALLY) UNDER AN AGREEMENT AND DECLARATION OF TRUST DATED JANUARY 2, 1926, AS AMENDED, WHICH IS HEREBY REFERRED TO, AND A COPY OF WHICH AS AMENDED HAS BEEN FILED WITH THE SECRETARY OF THE COMMONWEALTH OF MASSACHUSETTS. ANY AGREEMENT, OBLIGATION OR LIABILITY MADE, ENTERED INTO OR INCURRED BY OR ON BEHALF OF NEW ENGLAND ELECTRIC SYSTEM BINDS ONLY ITS TRUST ESTATE, AND NO SHAREHOLDER, DIRECTOR, TRUSTEE,

PHASE III - Developmental operation of the gasification process and integrated power plant

A utility is required as a cooperating party to carry out this program. In Phase I the utility will

supply technical information on its existing power plant - plant layouts, equipment drawings, design and operating data

provide price and availability information on oil fuels which might be utilized in the plant currently and/or in the future

give information concerning the generation requirements placed on the plant - daily and season variations in loading; availability of the plant for modification and development tests

cooperate in the selection of an engineering firm to prepare preliminary drawings, cost estimates and schedules for the installation of an oil gasifier/desulfurizer

assist in selecting various options in the design of the system - location of the equipment, choice of burners, means of by-pass, technique for temperature control, etc.

evaluate in cooperation with Westinghouse and EPA the effectiveness and economy of the oil gasifier/desulfurizer in power generation and pollution abatement

Westinghouse, under contract to EPA, will direct the work of the engineering firm as it prepares, under subcontract to Westinghouse, the preliminary gasifier/desulfurizer system designs and cost estimates for the utility power plant. EPA will provide general guidance and funding for Phase I.

New England Electric System (NEES) has been seeking a promising method to control stack emissions. We have studied this proposed oil gasification project and give it our strong support. Specifically, this will serve as a Letter of Intent to agree to participate in the program by providing the necessary technicial data and engineering manpower to complete the tasks and evaluations through Phase I. Two site options are suitable for the location of a demonstration plant - one at our Salem Harbor Station (#1), Salem Harbor, Mass. and one at our Manchester Street Station (#12) in Providence, R. I. We shall commit nine-man months of engineering (essentially the full time services of a senior engineer) to assist in the Phase I study.

At the conclusion of Phase I all participants in the program - NEES, Westinghouse and EPA - will mutually consider whether the program should be carried further. If technical and economic prospects of the installation appear favorable to our Management, we shall cooperate in making a joint proposal to EPA for completion of Phase II and III of the program.

Very truly yours

Harold Lurie

Harold Lurie
Director of Research & Development

m/

cc: R. P. Hangebrauck
I. W. Ellsworth

✓

APPENDIX L

Invitation to Bid

Westinghouse Electric Corporation

Research and Development Center

Beulah Road
Pittsburgh Pennsylvania 15235

(412) 256-7473

February 20, 1973

SUBJECT: INVITATION TO BID, NUMBER 6087

You are invited to submit a cost plus fixed fee proposal for Phase I, Attachment D, in enclosed document titled "Atmospheric Pressure Fluidized Bed Oil Gasification/Desulfurization Demonstration Plant." Contemplated Phases II and III, described also in the enclosed document, are not included in this solicitation.

Westinghouse Research and Development Center, as a prime contractor to the Environmental Protection Agency is studying a facility to convert high sulfur oil to clean gaseous fuel for an electrical power plant. The planned installation site is an existing power plant of the New England Electric System at Providence, R.I., where the fuel conversion system and combustion system would be retro-fitted to an existing convensional steam generator and turbine. The planned program covers three phases:

Phase 1	Design and Cost Studies
Phase 2	Installation of the Retro-Fitted fuel conversion and combustion System
Phase 3	Performance Studies and Experimentation

This solicitation 6087 invites proposals for Phase 1 only.

The work to be performed under Phase 1, Design and Cost studies consists of two parts:
(a) Feasibility, preliminary design, and cost studies.
(b) Preparation of a "Request for Proposals" package suitable for use in future solicitation of fixed price proposals for Phase 2.

Westinghouse approval will be required prior to proceeding with above Part (b) of Phase 1.

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The attachments hereto identify the precise nature of work to be performed under this program, together with background information now available on the nature of the proposed system.

Westinghouse requests that proposals (three copies please) in response to this solicitation include:

1. Identification and qualifications of Bidder's personnel who will have prime technical responsibility for the project and others who will perform a significant engineering work on the project; and
2. Identification of Bidder's previous experience on work related to the scope of work; and
3. A breakdown of estimated costs for Phase 1, showing materials, labor plus applicable rates, any other items of cost, and fee; and
4. A proposed schedule for completion of Phase 1, Part (a) and Part (b), based on subcontract award for Phase 1 in March, 1973.

Westinghouse requests submittal of your proposals by March 26, 1973. Prior to submittal of your proposals, Westinghouse further asks that you propose a time for a discussion meeting at Westinghouse Research Center. The purpose of such meeting will be to assure that all applicable aspects of this solicitation are understood; the meeting will be arranged individually with each Bidder on separate occasions.

Dr. D. L. Keairns, Manager, Fluidized Bed Research, Westinghouse Research Center, is responsible for the technical direction on this proposed subcontract; he may be reached by telephone on (412) 256-7345. R. A. Limerick, (412) 256-7473, is responsible for the administration.

Very truly yours,

R. A Limerick

R. A. Limerick
Senior Buyer

lv

Attachments

ATMOSPHERIC PRESSURE FLUIDIZED BED OIL GASIFICATION/DESULFURIZATION DEMONSTRATION PLANT

A demonstration plant program for the atmospheric pressure fluidized bed oil gasification process developed by Esso England has been established. The following documents provide

- review of the concept (Attachment A)
- summary of the demonstration plant program (Attachment B)
- data on the demonstration plant design (Attachment C)
- scope of work to be performed by an engineering contractor for Phase I (Attachment D)

The request for proposal covers Phase I of the program.

ATTACHMENT A

ATMOSPHERIC PRESSURE FLUID BED OIL GASIFICATION

The gasification/desulfurization of residual oil at atmospheric pressure has been developed by Esso England and evaluated by Westinghouse under contract to the Office of Research and Monitoring, ORM, of the Environmental Protection Agency. The concept is being studied for the purpose of producing, on-site, a low sulfur fuel gas suitable for power plant utilization in a conventional boiler. Esso Petroleum Company has provided experimental information on the oil gasification/desulfurization process based on small scale batch fluidized units and a 750 kW continuous unit. Westinghouse has carried out preliminary conceptual design studies to evaluate the commercial process and has carried out a preliminary evaluation of the demonstration plant facility.

GASIFICATION/DESULFURIZATION CONCEPTS

Two possible modes for gasification/desulfurization operation are the regenerative mode and the once-through mode. Fig. L-1 illustrates the major process streams and identifies the basic elements of the two operational modes.

The elements of the regenerative operation are the gasifier vessel and the regenerator vessel. The gasifier is an air-fluidized bed of lime operated at 1600°F with sub-stoichiometric air, $\sim$ 20% of stoichiometric. The oil is injected into the gasifier vessel where it cracks and is partially combusted to form a hot, low sulfur fuel gas. Hydrogen sulfide is produced during the gasification which reacts with the lime to produce a sulfided lime.

$$H_2S + CaO \rightarrow CaS + H_2O$$

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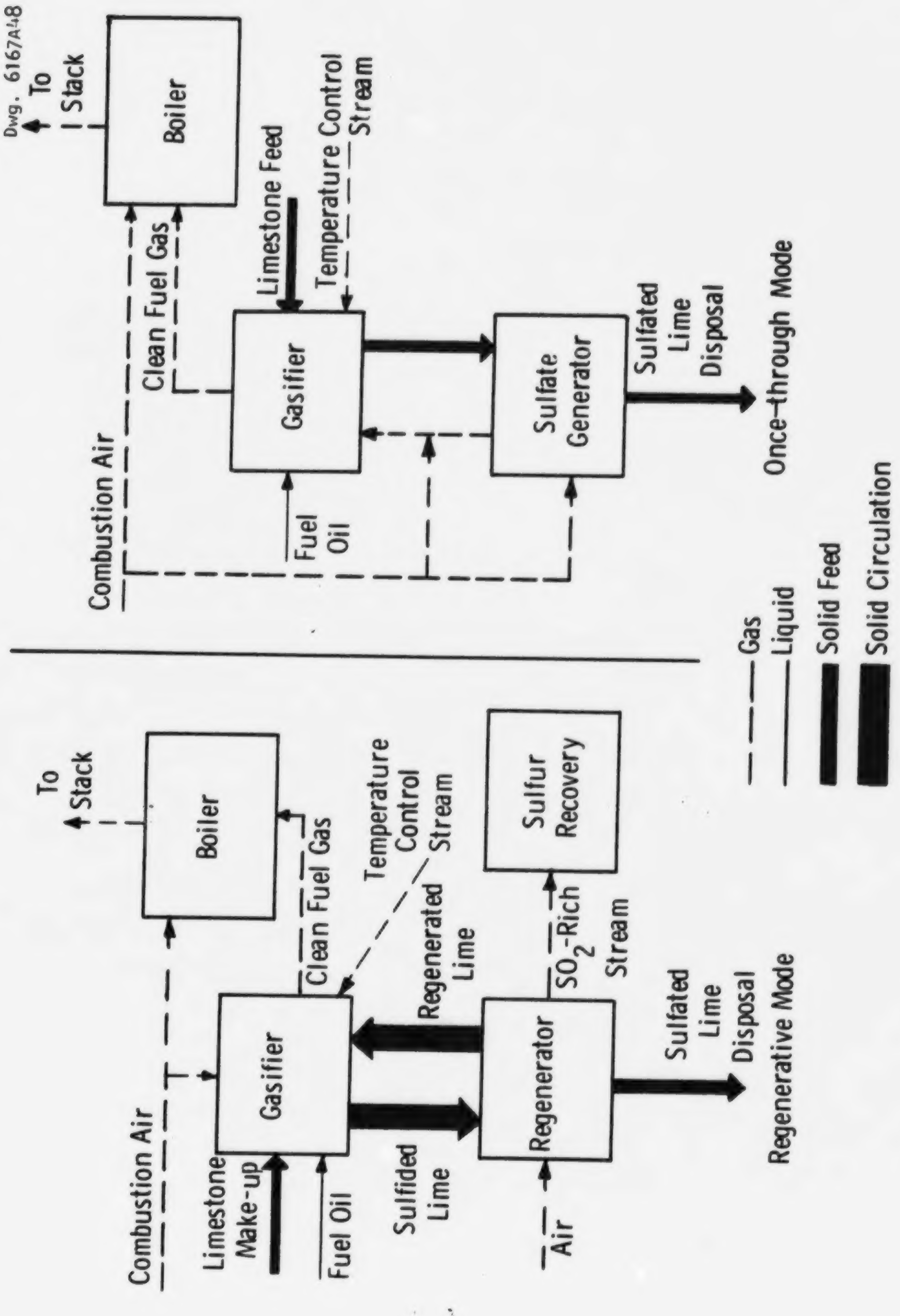

Figure L-1 Modes of Operation

The fuel gas is transported to the boiler burners where combustion is completed and the sulfided lime is sent to the regenerator. The regenerator is an air-fluidized vessel operated with a slight excess of air at 1900°F. Regeneration takes place by reaction of oxygen with the utilized lime to give an SO_2 rich stream (of about 10 mole % SO_2) and a regenerated lime having a slightly decreased activity compared to that of fresh lime.

$$CaS + \frac{3}{2} O_2 \rightarrow CaO + SO_2$$

The SO_2 stream is transported to a sulfur recovery system and the regenerated lime is returned to the gasifier along with a stoichiometric amount of fresh make-up limestone.

The elements of the once-through operation shown in Fig. L-1 consists of a gasifier vessel and a sulfate generator, or pre-disposal vessel. The operation of the gasifier for once-through limestone operation is the same as for the regenerative operation. The sulfate generator operates similarly to the regenerator but at a lower temperature (∿ 1500°F) so that the sulfided lime from the gasifier is converted to calcium sulfate rather than calcium oxide. The dry calcium sulfate may be disposed of while the gas stream from the sulfate generator is sent to the gasifier. A limestone addition rate of 3-4 times that used in the regenerative operation is necessary in the once-through operation to achieve similar sulfur removal of 90-95%.

EXPERIMENTAL WORK

Under contract to the Office of Research and Monitoring, Esso (England) has carried out laboratory tests on atmospheric-pressure batch fluidized bed equipment to investigate lime sulfur absorption and lime regeneration operating variables and pilot plant tests on a 1 MW continuous gasifier-regenerator system(1,2). The results of the Esso Petroleum Company experimental program have identified the critical phenomena associated with atmospheric-pressure gasification/desulfurization, and have established the probable operating conditions and behavior of a commercial system. This evaluation has been based on the results of the small scale batch work. The batch work has been carried out in conjunction with the construction and operation of the 1 MW continuous pilot unit.

DESIGN

Energy and material balances provide the basic information with which the feasibility of applying the gasification/desulfurization concept as an add-on to an existing boiler, or as a new plant design feature, have been examined. The feasibility of the retrofit concept has been examined in terms of the availability of space in an existing power plant, the modifications necessary to retrofit an existing boiler, and the performance of a modified boiler. The specifications assumed for the conceptual design are presented in Table 1.

TABLE 1

PROCESS SPECIFICATIONS

Sulfur Removal: 90-95%
Fuel Oil: 3 wt % sulfur; LHV = 17,700 Btu/lb
Limestone: 98.6% CaO yield

DESIGN VARIABLES

Boiler Size: 600 MW
Load Factor: 40%, 80%
Turn-Down: 4/1

OPERATING VARIABLES

	Regenerative System	Once-Through System
Gasifier temperature	1600°F	1600°F
Regenerator (sulfate generator) temperature	1900°F	1500°F
Stone make-up rate	1 mole CaO/mole sulfur	3 moles CaO/mole sulfur
Air/fuel ratio	20% of stoichiometric	20% of stoichiometric
Limestone utilization	∿ 5 wt % sulfur in bed	∿ 19 wt % sulfur in bed
Fluidization velocity	8 ft/sec	8 ft/sec
Minimum fluidization velocity	3 ft/sec	1 ft/sec
Particle sizes (avg.)	∿ 2000 μ	∿ 1000 μ
Gasifier bed depth (static)	2.5-3.5 ft	3.5-4.0 ft

Retrofit Concepts: The feasibility of converting an existing boiler to one which utilizes the fuel from a gasification/desulfurization system depends on a number of factors, many of which will differ from one boiler to the next. The space available for a gasification/desulfurization system in an existing power plant, the modifications necessary to retrofit an existing boiler, and the performance of a retrofit boiler will depend on the specific gasification/desulfurization system design and choice of operating conditions, the location of the gasification/desulfurization system in the plant, the type of boiler (coal-, oil-, or gas-fired), size of boiler, the turn-down needed, the boiler load factor, and the specific design features of the boiler.

The gasification system may be placed internal to the boiler (directly beneath) or external to the boiler. Preliminary considerations seem to favor the external retrofit design over the internal retrofit design, and the feasibility of retrofitting a coal-fired boiler over the feasibility of retrofitting an oil- or gas-fired boiler. The internal design concept appears to be limited by the available space beneath the boiler, and the cost for modifying the boiler. The external design concept should reduce modification costs and should allow greater uniformity and flexibility in system design. Coal-fired boilers have more potential space near the boiler than oil- or gas-fired boilers and may provide some of the needed solids handling and particulate clean-up equipment. Boiler derate should be of little concern with coal-fired boilers because they are designed to operate with slag on the boiler heat transfer surface. Derate may be a concern with oil- and gas-fired boilers due to a reduction in the flame temperature and a redistribution of the heat release within the boiler which may occur with the fuel.

In contract to boiler retrofit considerations, the feasibility of incorporating a gasification/desulfurization system into a new boiler design will be limited only by the overall economics of the system and the market potential for new oil-fired boilers. The total space occupied by the gasification/desulfurization system will be a small percentage of the total plant volume. Also, because of the flexibility in boiler design, boiler performance will not be affected by the presence of the gasification/desulfurization system.

Plant Layout: Fig. L-2 shows a plant layout for a 600 MW once-through gasification/desulfurization system. The plant consists of two gasification modules, each utilizing air from one of the two power plant forced draft fans present in the existing boiler.

EVALUATION

The design study, coupled with the experimental work of Esso Petroleum Company, points out the technical and economic feasibility of oil gasification/desulfurization as a retrofit SO_2 control system for utility boilers, or as an SO_2 control system to be incorporated into a new boiler design. A market for retrofit and new oil-fired boilers exists.(3)

Preliminary investigations indicate that, overall, the once-through operation may be somewhat more attractive to a utility customer than the regenerative operation. Capital investment is reduced with the once-through operation and, although the limestone feed rate is expected to be three times the rate with regenerative operation, the operating costs for once-through operation may be less than those for the regenerative operation. A complete cost breakdown for once-through and regenerative operation has been presented.(4,5) Once-through operation has fewer technical problems at this time, and is an overall simpler process than regenerative operation.

A comparison is made in Table 2 between atmospheric pressure oil gasification/desulfurization and the alternative schemes of low-sulfur oil and stack gas cleaning. Capital costs and operating costs are compared for new and retrofit systems. Oil gasification/desulfurization compares favorably with low-sulfur oil and stack gas cleaning, based on preliminary cost estimates. A reduction of about 40% in the capital costs involved in stack gas cleaning is estimated for new and retrofit gasification/desulfurization systems. Operating costs appear to be about

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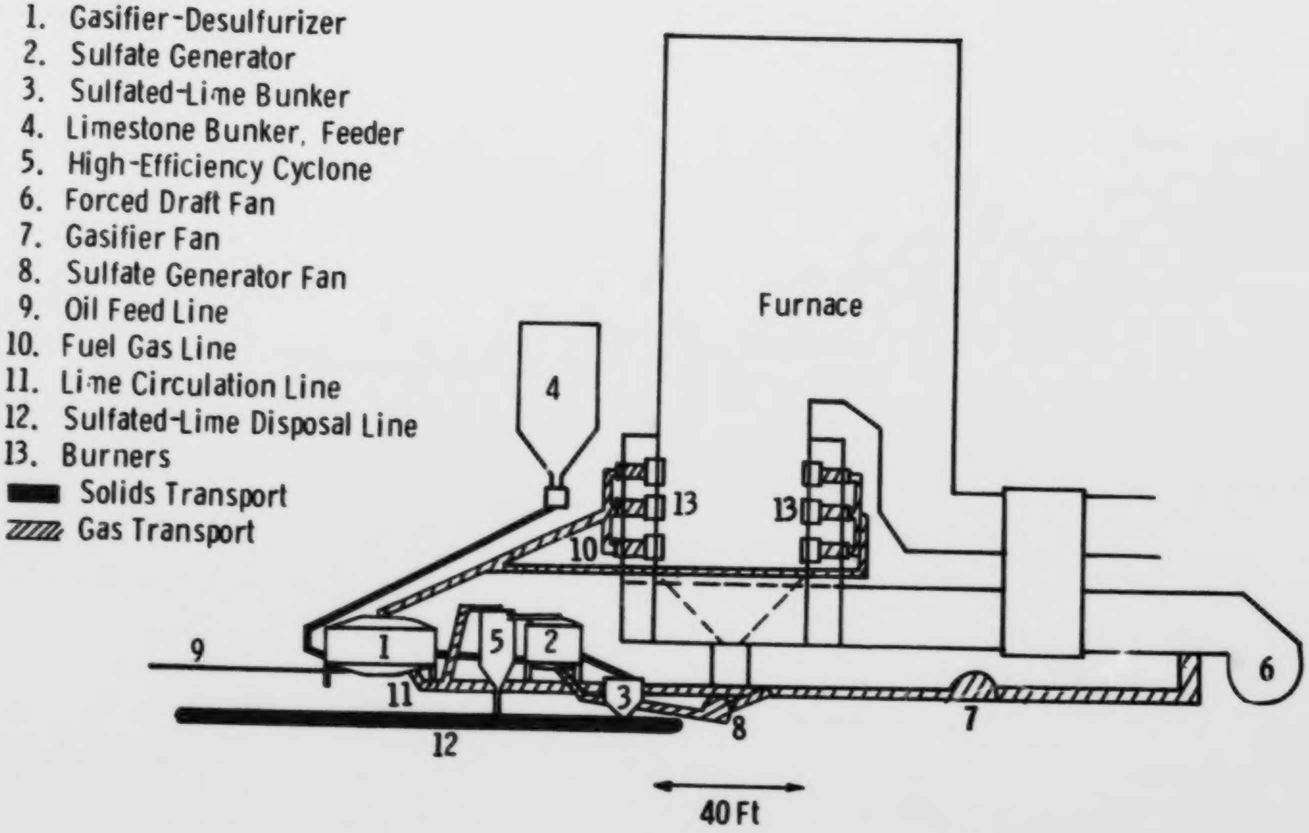

Figure L-2—600MW Boiler Retrofit Once-through Design

TABLE 2

ASSESSMENT OF FLUIDIZED BED OIL GASIFICATION/DESULFURIZATION

Basis: 3% Sulfur, 90% Sulfur Removal, 600 MW Capacity, 80% Load Factor

	Low-Sulfur Oil	Stack Gas Cleaning	Oil Gasification	
			Regenerative Operation	Once-Through Operation
COST				
Capital, \$/kW				
New	---	25-40	12-15	8-10
Retrofit	(a)	40-75	22-27	18-22
Fuel Adder, ¢/10^6 Btu				
New	25-35	11-14	9.5-16.0	9-10.5
Retrofit	25-35[a]	14-20	11-18	10.5-12.5
EFFICIENCY (THERMAL)	---	0.95-0.98	0.89-0.96[b]	0.96-0.97[b]
ENVIRONMENTAL FACTORS				
SO_2, (lb/10^6 Btu)	0.35	0.45	0.35	0.35
NO_x, (lb NO_2/10^6 Btu)	0.40	0.8	0.16	0.16
Particulates, (lb/10^6 Btu)	0.06	0.05	0.02-0.2[c]	0.02-0.2[c]
Solid Waste, ft^3/MW-day	---	25	15	45
S REMOVAL				
Stone	NA		Limestone	Limestone
Ca/S	NA		∿ 15[d]	3.0
Make-up Ca/S	NA		1.0	NA

[a] Equipment modifications are required when converting from gas or coal to low sulfur oil.

[b] Overall efficiency is dependent on mode of temperature control.

[c] 0.02 figure based on installing electrostatic precipitator (ESP).
0.2 figure based on installing high efficiency cyclone before burners and no ESP.

[d] Ca/S rate dependent on regenerator temperature control scheme.

the same for once-through stack gas cleaning and regenerative gasification/desulfurization with sulfur recovery. Once-through gasification/desulfurization may reduce the operating costs 30 to 50% as compared to stack gas cleaning. The cost estimate indicates that the operating cost with low-sulfur fuel oil will be about 30-50% greater than the operating cost for gasification/desulfurization. These conclusions are based on the desulfurization of high-sulfur residual oil (3 wt % sulfur) and may be altered when a lower sulfur oil is considered (1 to 1.5 wt % sulfur). Environmental factors are also compared in Table 2. The low-sulfur oil is advantageous in that capital costs are limited to possible boiler modifications necessary when changing from gas or coal to low-sulfur oil. On the other hand, operating costs are higher than those for stack gas cleaning or oil gasification/desulfurization. Capital costs are extremely high with stack gas cleaning, especially on the retrofit case, while operating costs are very near those estimated for oil gasification/desulfurization.

Table 3 lists some advantages of atmospheric-pressure oil gasification over stack gas wet scrubbers.

TABLE 3

ADVANTAGES OVER STACK GAS WET SCRUBBERS

- Corrosion and fouling problems minimized in SO_2 removal process and boiler (minimum SO_x and V)
- No flue gas reheat required
- Uses crushed limestone - no limestone pulverizing system needed
- Simplified disposal - dry solids and no disposal pond
- More compact system
- Reduced structural costs
- Lower auxiliary power requirement
- Reduced energy cost
- Improved NO_x control
- Potential market for spent CaO

CONCLUSIONS

PERFORMANCE

- The concept has been technically demonstrated with a 750 kW development gasification plant.
- Sulfur removal up to 95% can be achieved.
- Nitrogen oxide emissions of 150 ppm appear possible.
- Particulate emissions will be higher than conventional oil- or gas-fired systems but can easily be removed to achieve proposed standards.
- Further development effort is required in key areas -- e.g., calcium sulfate generation for once-through operation, temperature control, sulfur recovery.

ECONOMICS FOR COMPARABLE POLLUTION ABATEMENT

- Capital cost of a retrofit, once-through gasification system may be 50 to 70% less than a retrofit wet scrubbing system.
- Fuel adder cost for a retrofit, once-through gasification system may be 30 to 50% less than wet scrubbing, low-sulfur oil, or desulfurized oil.

MARKET

- Initial market is expected to be small boilers (< 600 MW) on the East Coast, in the Southwest where gas may be limited, or on the West Coast. Once-through operation may be favored over operation with sulfur recovery for these plants.

REFERENCES

1. Craig, J. W. T., G. L. Johnes, G. Moss, and J. H. Taylor; "Study of Chemically Active Fluid Bed Gasifier for Reduction of Sulfur Oxide Emissions", Interim Report, APCO Contract CPA 70-46, Esso Research Centre, Abingdon, Berkshire, England, August 1970.

2. Craig, J. W. T., G. L. Johnes, G. Moss, J. H. Taylor and D. E. Tisdall; "Study of Chemically Active Fluid Bed Gasifier for Reduction of Sulfur Oxide Emissions", Final Report, OAP Contract CPA 70-46, Esso Research Centre, Abingdon, Berkshire, England, June 1972.

3. Hauser, L. G.; "Optimum Uses of Energy Sources", paper presented at Spring Conference, Southeastern Electric Exchange, New Orleans, La., April 1971.

4. Newby, R. A., D. L. Keairns, and D. H. Archer; "Assessment of Fluidized Bed Oil Gasification for Power Generation", paper presented at the 65th Annual Meeting of the Air Pollution Control Assoc., Miami Beach, Florida, June 1972.

5. "Evaluation of the Fluidized Bed Combustion Process", Final Report, OAP Contract CPA 70-9, Westinghouse Research Laboratories, November 1971.

ATTACHMENT B

DEMONSTRATION PLANT PROGRAM

A three phase demonstration plant program has been conceived:

Phase I - Preliminary design and cost estimate of demonstration plant installed on an existing boiler plus the subsequent preparation of a bid package and a definitive estimate for Phase II work.

Phase II - Detailed design and construction of demonstration oil gasification process.

Phase III - Developmental operation of the gasification process and integrated power plant.

Westinghouse, as prime contractor to EPA, will subcontract for the engineering services for Phase I. Phase I has two sequential parts: 1) a preliminary design and budget estimate of investment for evaluation of the preliminary design, followed by decision to stop or proceed; 2) preparation of a bid package and a definitive installed cost estimate. The bid package is to be adequate to obtain comparable lump sum bids for the Phase II work. New England Electric System, as the cooperating utility will supply technical information on the Manchester Street power plant, provide price and availability information on oil fuels which might be utilized in the plant currently and/or in the future, supply information concerning the load requirements placed on the plant, cooperate in the selection of an engineering firm to prepare the preliminary design and bid package, assist in selecting various options in the design of the system and evaluate, in cooperation with Westinghouse, Esso (England) and EPA, the effectiveness and economy of the oil gasifier/desulfurizer in power generation and pollution abatement.

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The design effort will be based on experimental data from Esso (England) on their one MW continuous pilot plant, the conceptual design and assessment by Westinghouse and operating and environmental constraints established by NEES and local regulations. EPA will provide general guidance and funding for Phase I.

If the preliminary design confirms the effectiveness and economics of the system, a proposal will be prepared for Phases II and III. In Phase II the utility would work closely with an engineering firm in the design and installation of the gasifier/desulfurizer system and share in the costs of the design and installation.

In Phase III the plant would be operated to collect basic data on the operation of the gasifier/desulfurizer system, data would be interpreted and analysed and an evaluation of the effectiveness, technical and economic, of the process in power generation and pollution abatement would be made.

ATTACHMENT C

PRELIMINARY ASSESSMENT OF PROPOSED MANCHESTER STREET DEMONSTRATION PLANT

Design Considerations

Plant Operation Considerations

Experimental Program

Assessment of Manchester Street Site
- Site Plan & Elevation
- Basis for Assessment
- Projected Solids Disposal
- Material Balance
- Preliminary Design
- Cost Estimate
- Conceptual Plant Layout

DESIGN CONSIDERATIONS

- Process options

  Two potential locations (see plant diagrams)

  Temperature control method

  Water injection
  Steam injection
  Stack gas recycle

  Once-through operation or regenerative operation

- Burner and fuel piping

  Preliminary design indicates:

Case (a)	Case (b)
. 56 x 66 in. required for 110 x 10^6 Btu/hr burner.	. 70 in. diameter port for 250 x 10^6 Btu/hr burner.

- Limestone handling

  Are the coal bunkers available?

  Receiving facility?

- Access for installation

- Air blower

  Space requirement for separate blower and ducting

- Sulfur recovery if regenerate stone

- Height

  Does not appear to be a limitation

  How critical is the indicated height?

- Particulate emissions

  Requirement

  Present particulate removal equipment

- Solids disposal

- Operating requirements

PLANT OPERATION CONSIDERATIONS

- Process variables
    - Fluidization velocities
    - Bed depths
    - Sorbent particle sizes
    - Air/fuel ratio
    - Limestone make-up rate
    - Bed Temperatures
- Fuel
    - Dual fuel capability
    - Oil characteristics
- Sorbent
    - Limestone/dolomite
    - Regenerative/once-through
- Availability of plant for experimental operation
- Plant operation
    - Capacity factor
    - Load range

EXPERIMENTAL PROGRAM

The demonstration plant should be equipped to monitor the conditions of major streams and vessels to demonstrate the overall performance of the atmospheric pressure fluidized bed gasification-combustion system. The following areas should be emphasized in investigations with regenerative or once-through operation:

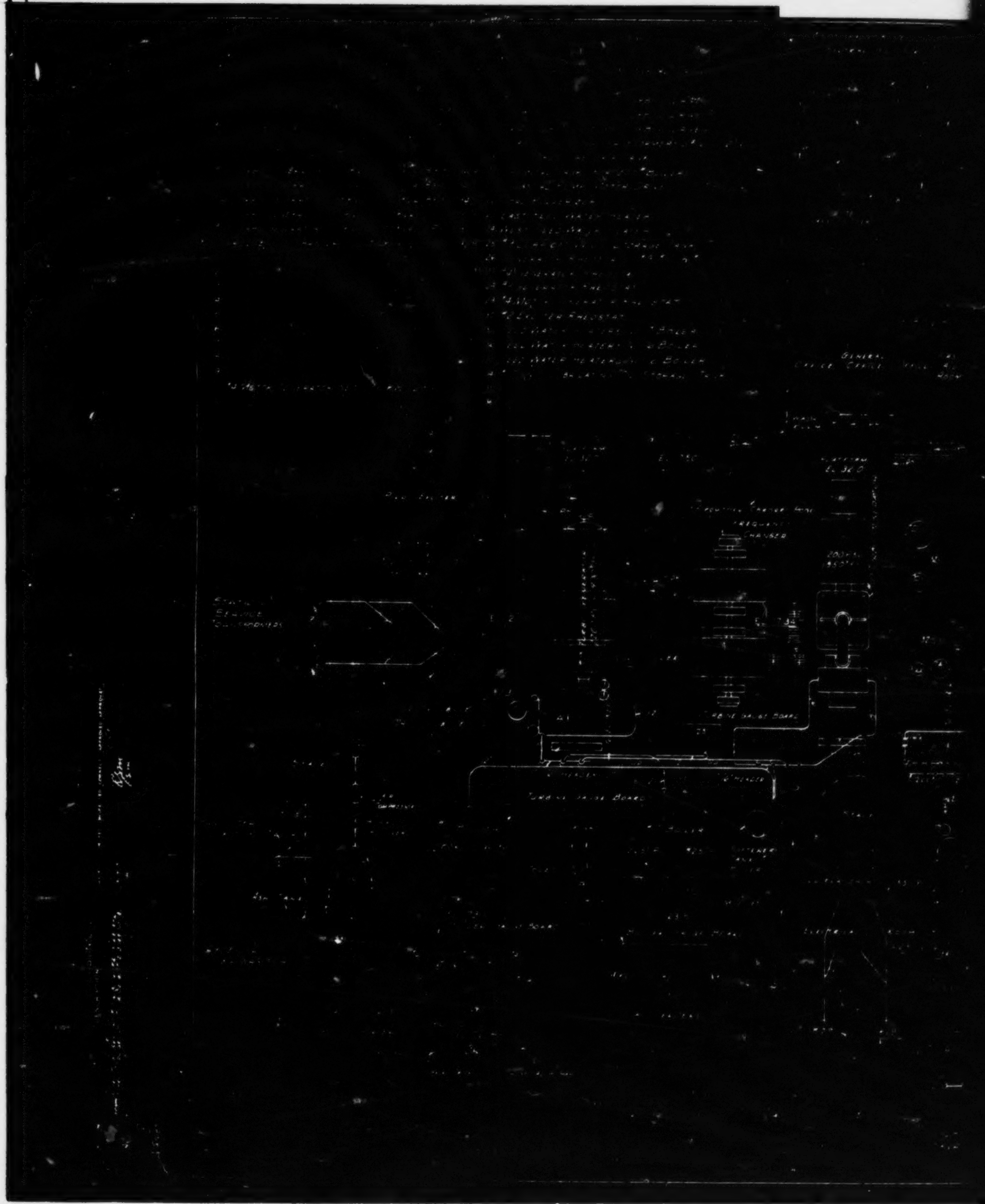
- Boiler performance
    - Steam conditions
    - Combustion characteristics of flame
    - Long term erosion/corrosion of boiler and burners
- Gasification/desulfurization system performance
    - Performance of temperature control schemes
    - Gasifier thermal efficiency
    - Fuel and solids distribution in gasifier
    - Sulfur recovery-regeneration performance (regenerative mode)
    - Sulfate generator performance (once-through mode)
    - Solids circulation and control
    - Long-term erosion/corrosion, material build-up in system
- System performance
    - Overall system efficiency
    - Optimum operating conditions for a given sulfur removal (in terms of limestone make-up rate, bed depth, particle size)
    - Pollution control - NO_x, SO_2, particulates, metal retention
    - Effects of turn-down on system performance
    - Overall system operation

The experimental program will be determined on the basis of the concept and the availability of the plant for experimental work.

ASSESSMENT OF MANCHESTER STREET SITE

Site Plan & Elevation

I A


GENERAL
PLATFORM
EL 32.0
FREQUENCY CHANGER
STATION SERVICE TRANSFORMERS
TURBINE GAUGE BOARD
HEADER
SOFTENER

B

Figure L-3—Manchester Street Station

Operating Floor Plan

/

A

B

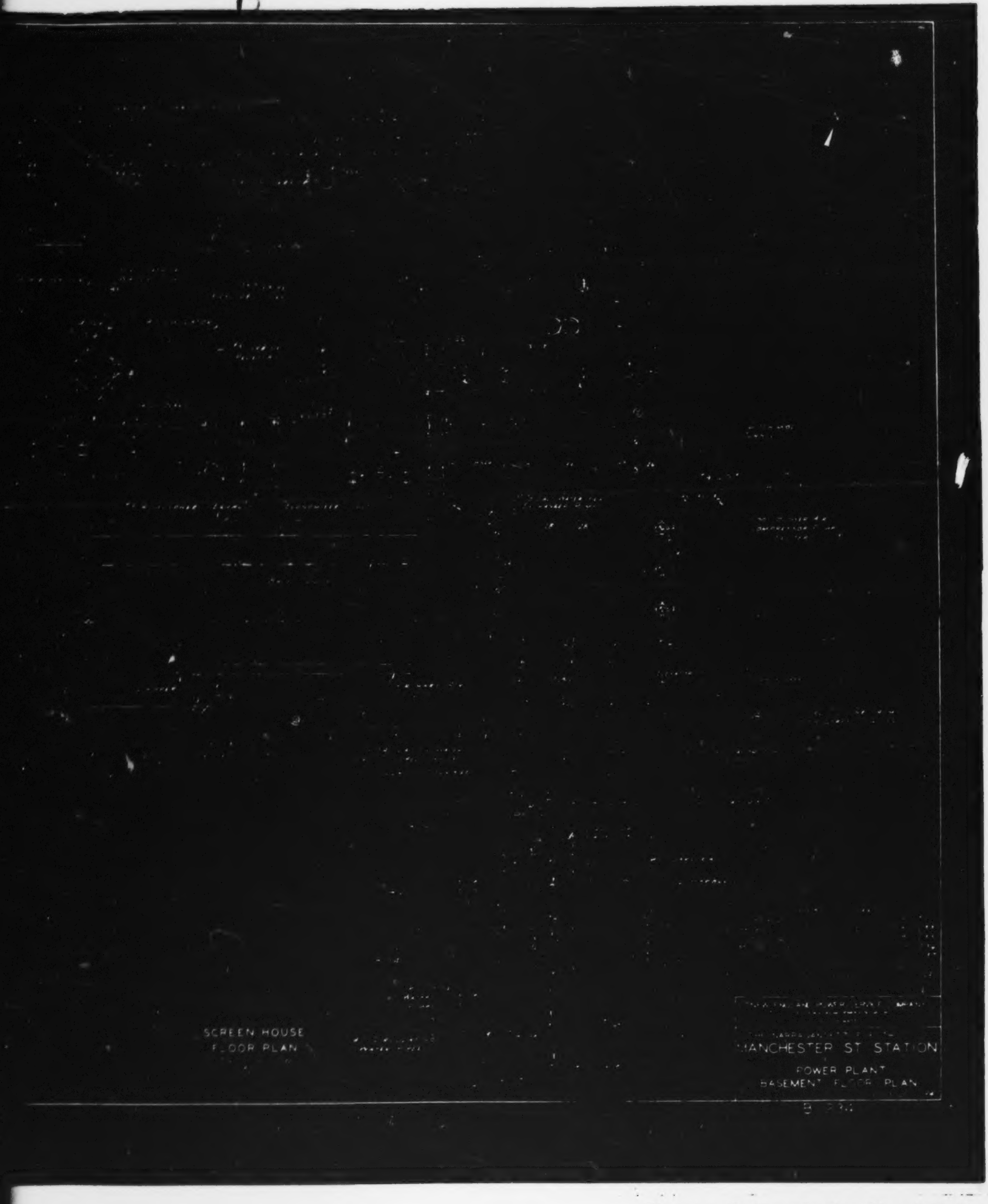


Figure L-4—Manchester Street Station Basement Floor Plan

L-27

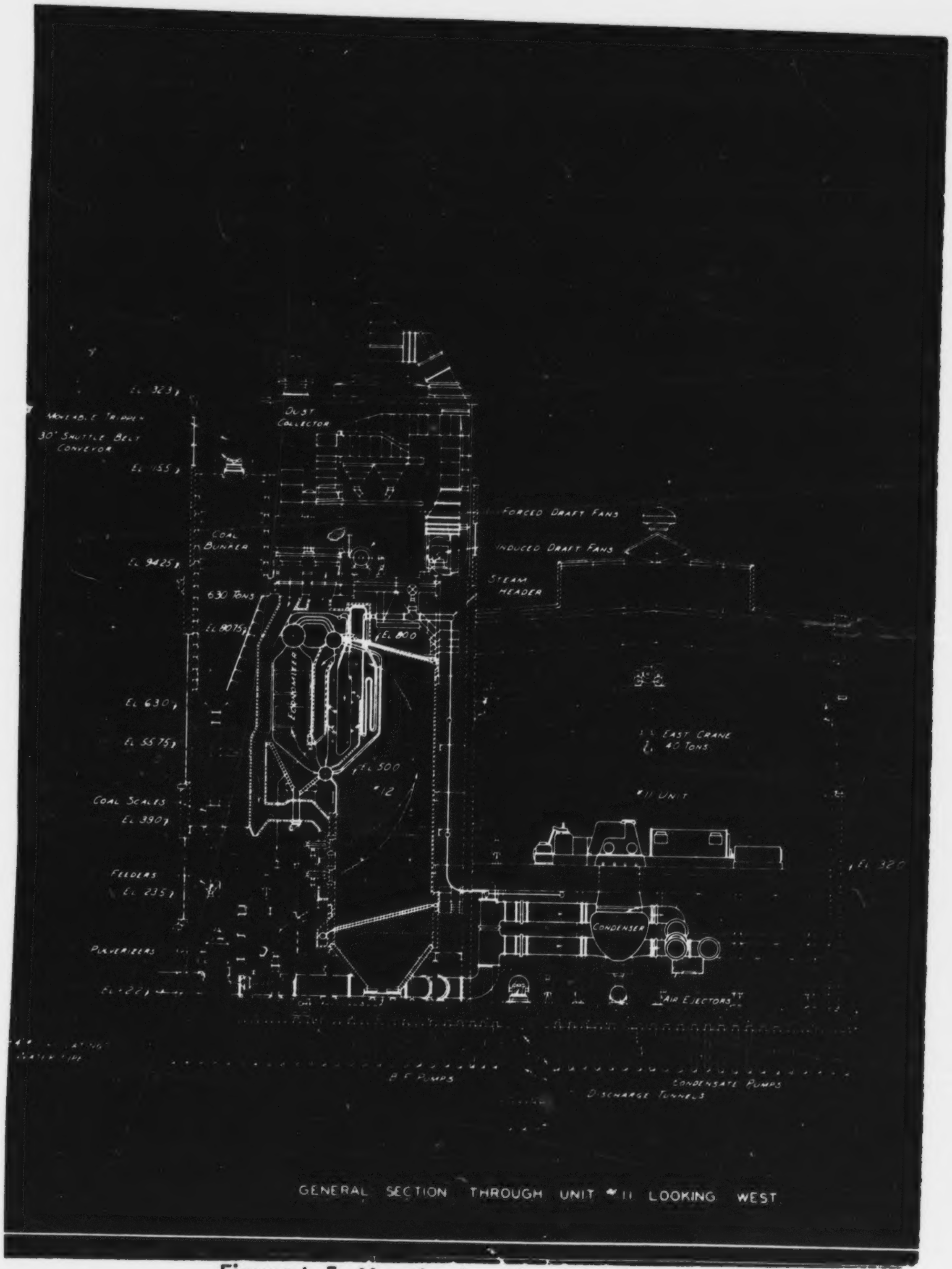


Figure L-5—Manchester Street Station
Basement Floor Plan

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Basis for Assessment

Station	Manchester
Unit	#12
Capacity, MW	46
Heat rate, Btu/kWh	10,400
Gasifier	
Fluidizing velocity	6 fps
Particle size	< 2000μ
Air/fuel ratio	20% of stoichiometric
Bed depth	2.5 ft
Bed temperature	1600°F
Thermal efficiency	95%
Sulfur removal	90-95%
Limestone feed rate	1 mole Ca/mole sulfur (regenerative) 3 moles Ca/mole sulfur (once-through)
Regenerator	10% SO_2 product gas; 2000°F
Sulfate generator	Complete conversion of CaS to $CaSO_4$; 1500°F
Temperature control	Water injection or stack gas recycle
Solids for disposal	0.6 $(yd)^3$/MW-day
Fuel properties	3 wt % sulfur; 18,000 Btu/lb LHV
Limestone properties	95% $CaCO_3$

PROJECTED SOLIDS DISPOSAL

Boiler	% Sulfur Oil	Plant Capacity Factor	% S Removal	Solids Disposal yd^3/day Regenerative Operation	Solids Disposal yd^3/day Once-Through Operation
Manchester St. #12; 50 MW	3	1.0	90[(a)]	< 30	< 90
		0.7	90[(a)]	< 20	< 60
		0.7	80[(b)]	10-15	30-40
	2	1.0	90[(c)]	< 20	< 60
		0.7	90[(c)]	< 14	< 40
		0.7	65[(b)]	< 10	∿ 20

(a) Equivalent to 0.33 lb $SO_2/10^6$ Btu emission; greater than 90% removal has been demonstrated.

(b) Would meet EPA new source emission std. of 0.8 lb $SO_2/10^6$ Btu.

(c) Equivalent to 0.22 lb $SO_2/10^6$ Btu emission.

Material Balance

	Regenerative Process	Once-Through Process
Fuel oil, lb/hr	28,000	28,000
Limestone, lb/hr	2,660	7,420
Lime ($CaSO_4$) disposal, lb/hr	1,430	5,200
Temperature control stream		
Water injection, lb/hr	10,400	10,400
Stack gas recycle, lb/hr	47,600	47,600
Sulfur recovered, lb/hr	800	---
Gasifier air, lb/hr	77,300	82,300

Preliminary Design

Plant Equipment

Equipment in space allocated

- gasifier
- sulfate generator
- regenerator (would not be needed if once-through operation)
- particulate removal - cyclones
- transport lines
  - oil, limestone, fuel gas, spent stone, water

External equipment

- limestone handling
  - receiving facility
  - storage
  - bunker
  - conveyors
  - feeders
- spent stone
  - storage
  - disposal
  - sulfur recovery -- if regeneration
- final particulate clean-up
- fan for gasifier air supply, (regenerator), sulfate generator,
  - pneumatic transport
- boiler
  - burner modification/boiler structure

Primary Equipment Design

Gasifier -

Cross-sectional area = 306 ft^2
Bed Depth = 2.5 ft
Freeboard = 7 ft

Sulfate Generator -

Cross-sectional area = 63.5 ft^2
Bed Depth = 3 ft
Freeboard = 6 ft

Regenerator -

Cross-sectional area = 23 ft^2
Bed Depth = 3 ft
Freeboard = 6 ft

Hot Gas Piping -

Inner Diameter = 2.5 ft

Burner -

Boiler	12
Capacity, Btu/hr	480×10^6
Existing Burners	6
Capacity/Burner, Btu/hr	80×10^6
High Temperature, Low Btu Burner Port Size (dual fuel -- low Btu gas and oil)	48" diameter max.
Outer Diameter of Port Block (if needed)	60" diameter
Feasibility of using existing burner ports based on preliminary contract with vendor	Yes

ATMOSPHERIC-PRESSURE FLUIDIZED BED OIL GASIFICATION DEMONSTRATION PLANT

COST ESTIMATE
(Installed)

Station	Manchester Street
Boiler	12
Capacity	50 MW
Oil Gasification System Cost	
Regenerative with sulfur recovery	$\$3.5 \times 10^6$
Once-through limestone throw-away	$\$2.4 \times 10^6$

Equipment Included in Cost Estimate

Gasifier

Regenerator (for regenerative system only)

Sulfur Recovery System (for regenerative system only)

Sulfate Generator

Particulate Removal Before Boiler

High Temperature Gas Piping

Burner Replacement

Limestone Receiving Facility, Conveyors and Feeders
(assumes use of existing coal bunkers for limestone)

Spent Stone Transport to On-site Storage and Storage Bin

Fan for Gasification System Air Supply

Experimental Instrumentation and Control

Oil Supply System (storage assumed available)

CONCEPTUAL PLANT LAYOUT

(For one or two possible locations)

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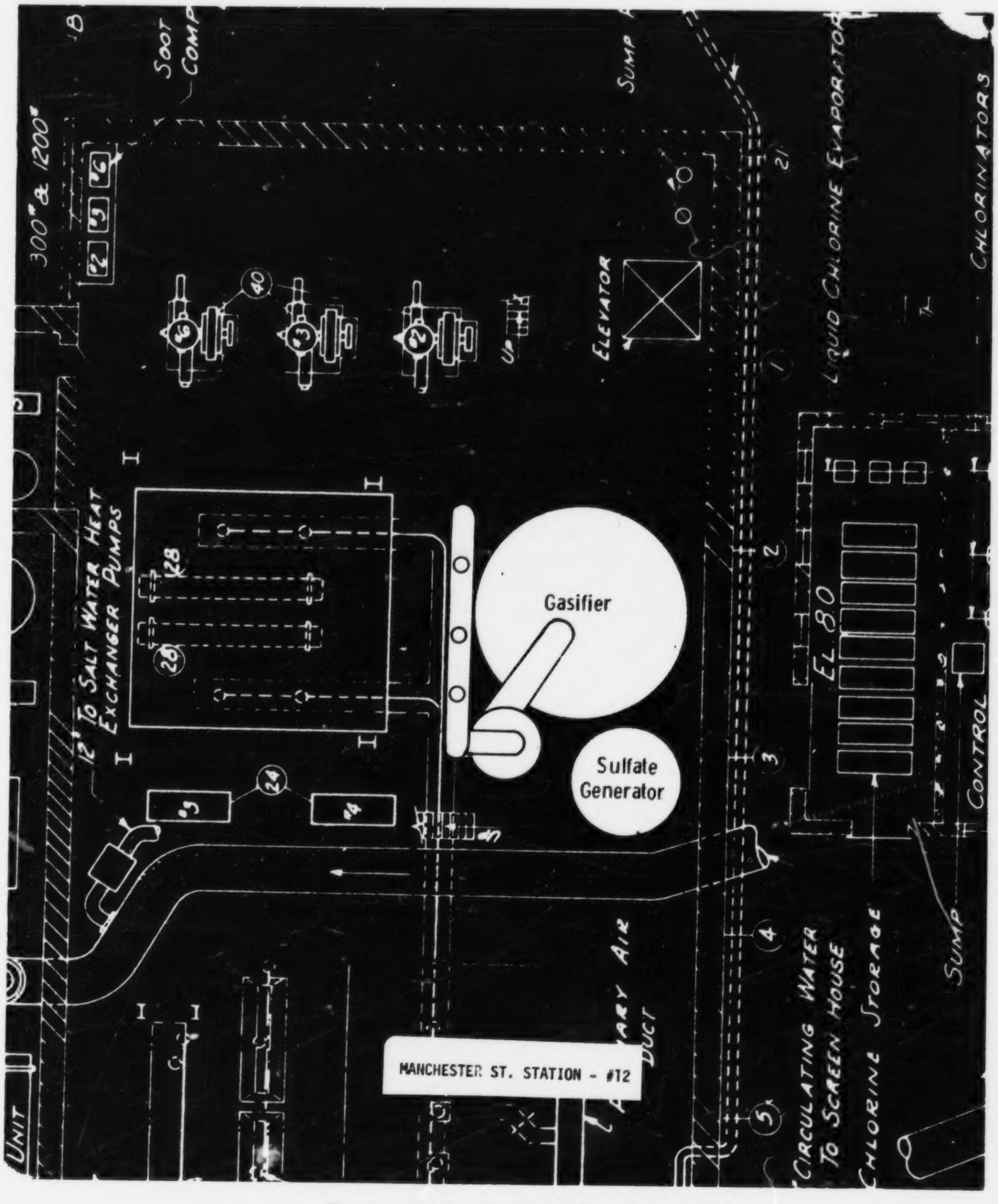

Figure L-6—Plan View of Gasifier Installation

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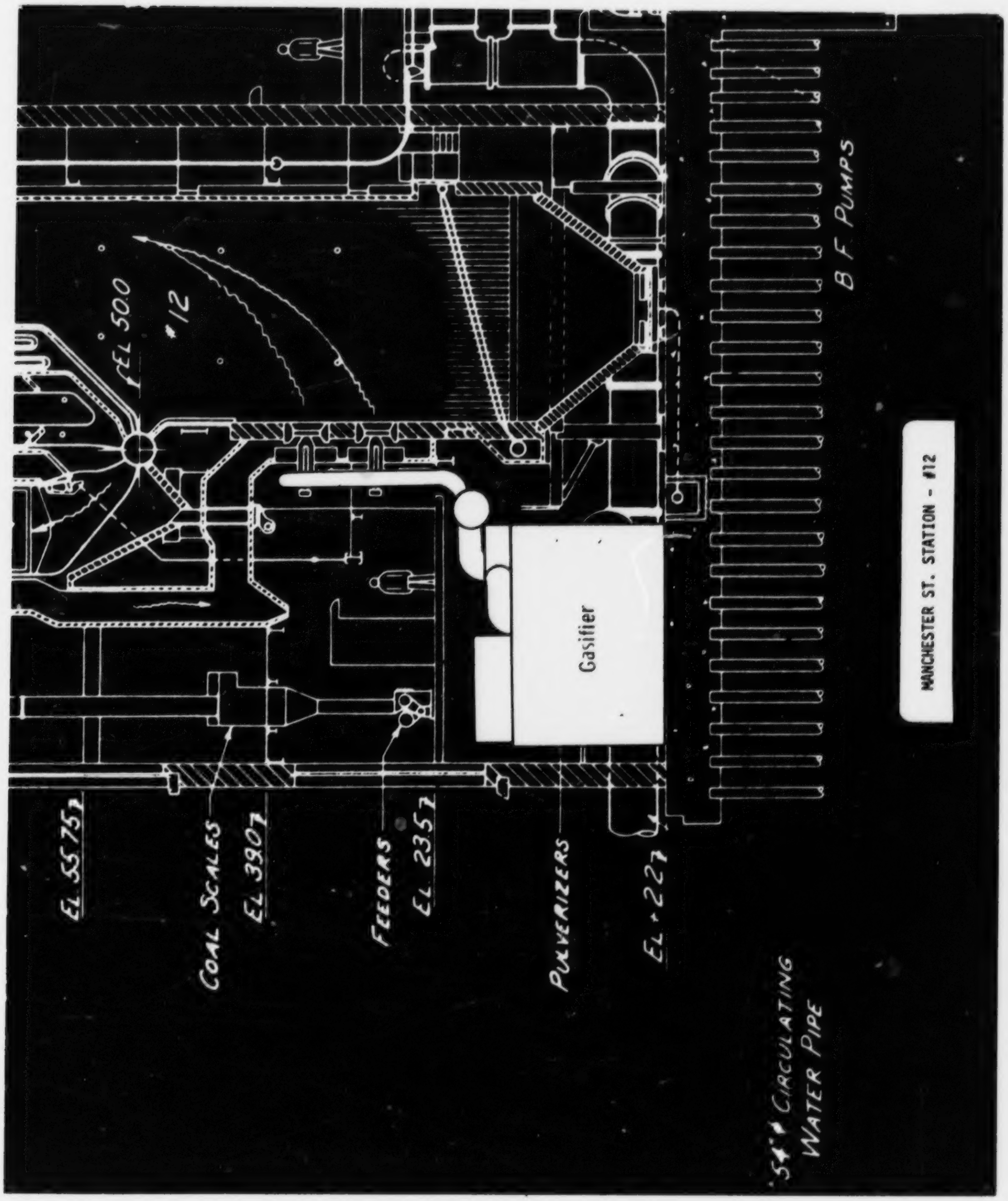

Figure L-7—Elevation View of Gasifier Installation

ATTACHMENT D

PHASE I - SCOPE

A. Goals

1. Preliminary design; performance, cost, and schedule estimates for funding and support building.
2. Design - bid package sufficient to obtain fixed price bid on detailed design and construction.
    - (a) Cost estimate and schedule for detail design, construction and operation.
    - (b) Experimental program for operational planning.
    - (c) Performance projection.
    - (d) Demonstration of process options selected on Esso pilot facilities -- where possible.

B. Scope of Work

Engineering Designs

- Review of Manchester Street Station site
- Process and engineering flow diagrams
    - Regeneration/once-through
    - Unit design/modular
    - Dual fuel capability
    - Solids processing for disposal
    - Alternative disposal techniques and economics
    - Plant modifications -- boiler, air supply, etc.
    - Process control -- temperature, load follow, start-up , shut down, emergency
    - Fuel oil and dolomite
- Plot plans
- Perspective drawings

- Layout
- Noise level evaluation
- Utilities distribution plan
- Equipment item list and specifications
- General description of facility
- General specifications (elect., inst., structural, etc.)
- Cost for detail design
- Cost for construction
- Experimental programs
- Cost of experimental program
- Statement of contractors responsibilities
- Statement of clients responsibilities

APPENDIX M

PARTIAL OXIDATION PROCESSES

A summary of partial oxidation processes for petroleum is enclosed.[1,2,3,5] The Shell and Texaco processes are the only partial oxidation processes commercially available which handle heavy fuel oil at pressure. The Koppers Totzec process can use heavy fuel at atmospheric pressure. Segas and Onia-Gegi are cyclic, catalytic processes which can process heavy fuel oil at atmospheric pressure. They produce ∿ 500 Btu/scf gas. None of these processes removes sulfur during the partial oxidation. Sulfur is removed by cooling the gas. Shell believes the vanadium and nickel content in the product gas leaving the gasifier is ∿ 50% of that fed.[3] All the vanadium, nickel, and Na are removed in the scrubbing system according to Texaco.[4]

The only known large scale commercial operations with air include[2,3,4]

- A Shell gasification system installed in 1971 to provide reducing gas for the Falconbridge Nickel Company's metallurgical operations in the Dominican Republic. The plant uses 12 Shell gasifiers operating on naphtha and air.
- A Texaco gasifier operated by Olin Mathieson Chemical in Morgantown, W. Va. in the late 1950s. The feedstock was anthracite coal fines.
- A Texaco gasifier operated, by ICI in England, on fuel oil and air for one year.

TABLE M-1

PARTIAL COMBUSTION PROCESSES

	Dayton Oil Gas Gas Light & Coke Co. GEIM	Koppers-Hasche	Distrigaz[†] Onia Autothermic P2	Koppers Totzec	Shell Texaco Montecatini Didier	Esso CAFB	Pressurized Fluidized Bed Oil Gasification
Process Feedstock	distillate	hydrocarbon gases	light distillate methane refinery gas LPG	heavy fuel oil	any gaseous or liquid hydrocarbon[*]	any oil	any oil
Catalyst	no	no	yes	no	no	no	no
Oxygen source	air	air	air/O_2	O_2	O_2/air	air	air
Steam	no	no	yes	yes	yes	perhaps	perhaps
Temperature, °F	1292-1562	1292-1562	?			1652	1652
Pressure, atm	1	1	1/12	~ 1	10-40	1	10-30
Tar	5-10% vol	---	---	?		---	---
Heating Value, Btu/scf	300-500	?	160-180/260	~ 300	/120	250-500	250-500
Efficiency (gas), %	79-82	95	89-92	70-80	80-85% (70%**)	94-97	94-97

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[*] Includes heavy oils, high sulfur, vanadium up to 500 ppm, nickel up to 100 ppm.

[**] Allows for oxygen plant, etc.

[†] Topse - S.B.A. Autothermic Process and Didier Catalytic Partial Oxidation Process are pressurized versions of this process. The process is partial oxidation followed by catalytic reforming to produce gas for ammonia synthesis.

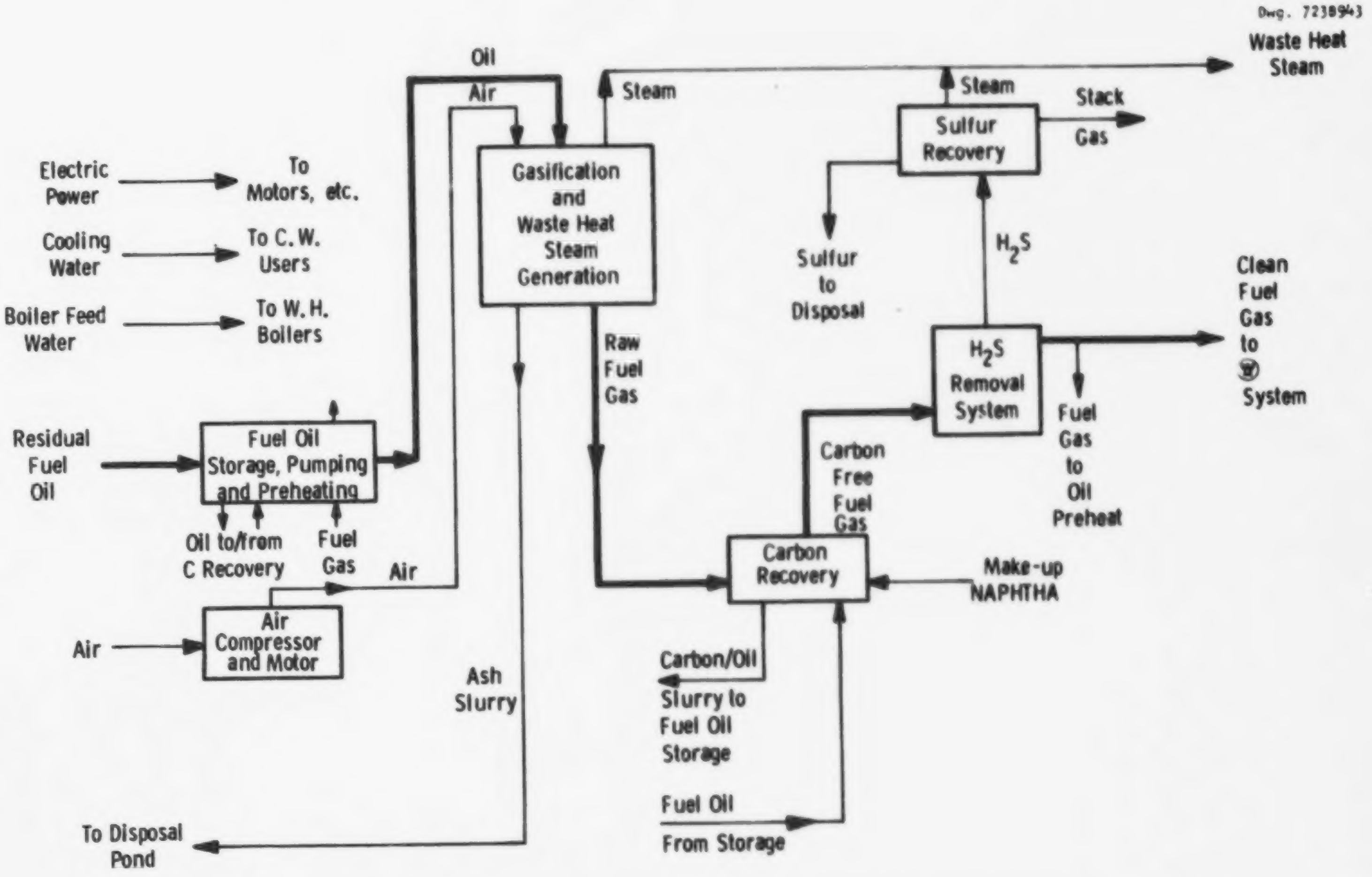

Figure M-1—Pressurized Oil Gasification for Gas Turbine Fuel - Shell and Texaco Processes

REFERENCES

1. Gasmaking, BP (1965).

2. "Oil Gasification for the Production of Methane, Hydrogen and Other Fuel Gases: An Engineering and Economic Analysis," Vol. III, SRI Project ECC-1203, August 1972.

3. Keairns, D. L. and D. H. Archer, Trip Report FBCOM 72-2, May 1972.

4. Discussions with Texaco on May 24, 1972.

5. Newby, R. A., D. L. Keairns and D. H. Archer. "Assessment of Fluidized Bed Oil Gasification for Power Generation." JAPCA, 23 (1), 23-29, 1973.

   Newby, R. A., D. L. Keairns, E. J. Vidt, D. H. Archer, and N. E. Weeks. "Fluidized Bed Oil Gasification for Clean Power Generation -- Atmospheric and Pressurized Operation." Paper presented at Third International Conference on Fluidized Bed Combustion, Hueston Woods, Ohio, Oct. 1972.

APPENDIX N

MATERIAL AND ENERGY BALANCES

Data at elevated pressures of 10 to 20 atmospheres* for the reactions taking place in the fluidized bed gasifier, regenerator, or sulfate generator do not exist, so the preliminary design is based on the atmospheric pressure results of Esso (England), with some modification. The following reactions are of importance:

1) Residual oil $\xrightarrow[\text{cracking}]{\text{thermal}}$ C(solid) + H_2 + H_2S + CH_4 + ...

2) $C + 1/2\ O_2 \rightleftarrows CO$

3) $C + O_2 \rightleftarrows CO_2$

4) $CO_2 + H_2 \rightleftarrows CO + H_2O$

5) $CaCO_3 \rightleftarrows CaO + CO_2$

6) $CaO + H_2S \rightleftarrows CaS + H_2O$

7) $CaCO_3 + H_2S \rightleftarrows CaS + H_2O + CO_2$

8) $CaS + 3/2\ O_2 \rightleftarrows CaO + SO_2$

9) $CaS + 2\ O_2 \rightleftarrows CaSO_4$

10) $CaS + H_2O + CO_2 \rightleftarrows CaCO_3 + H_2S$.

Reaction 1, thermal cracking of residual oils, will be altered by increased pressure, but the extent of alteration is not known. The effect of increased pressure on coking, or carbon deposition, taking place during cracking may be a critical factor in the pressurized operation. The distribution of hydrocarbons in the gaseous product and the

---

*Pressures up to 20 atmospheres are considered to allow for pressure drop in the gasification system prior to the gas turbine.

CO/CO_2 ratio in the product will differ from atmospheric-pressure results through the action of increased pressure on reactions 1 through 4. As a preliminary approximation, it is assumed that the gasifier at pressure will have the same temperature control requirements (same heat generation rate) as at atmospheric pressure. With this assumption many of the energy and material balances carried out for atmospheric-pressure operation apply to the pressurized case. This assumption must be tested experimentally, but it is a reasonable first approximation.

The calcination of limestone, reaction 5, is hindered by increased pressure and will be negligible at 20 atmospheres. This fact will slightly change the gasifier temperature control requirements from those at atmospheric pressure because of endothermic nature of reaction 5. Either of the two sulfur removal reactions, reactions 5 and 7, may apply to the process, depending on the lime regeneration scheme and the mode of operation (regenerative or once-through). Thermodynamically, both reactions 6 and 7 are high in sulfur removal potential at pressures up to 20 atmospheres. The equilibrium of reaction 7 is affected by pressure, while the equilibrium of reaction 6 is independent of pressure. Kinetically, Squires has noted that reaction 7 is faster than reaction 6 at pressure.[1] Thus, removal of sulfur is expected to be similar to that found for the atmospheric-pressure case, with the same operating temperature in the gasifier (∿ 1600°F) being suitable for either regenerative or once-through operation.

Figures N-1 and N-2 examine the thermodynamic behavior of the regenerator and sulfate generator as a function of temperature for pressures of 1, 10, and 20 atmospheres. Figure N-1 shows the mole fraction of SO_2 in the regenerator or sulfate generator product gas as is produced by reactions 8 and 9. Figure N-2 shows the fractional conversion of CaS to CaO and $CaSO_4$ in the regenerator or sulfate generator solid product. The

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[1] Ruth, L. A., A. M. Squires, R. A. Graff. "Desulfurization of Fuels with Calcined Dolomite. III. First Results for Reaction of $CaCO_3$ with H_2S." Paper presented at Los Angeles Meeting of American Chemical Society, March, 1971.

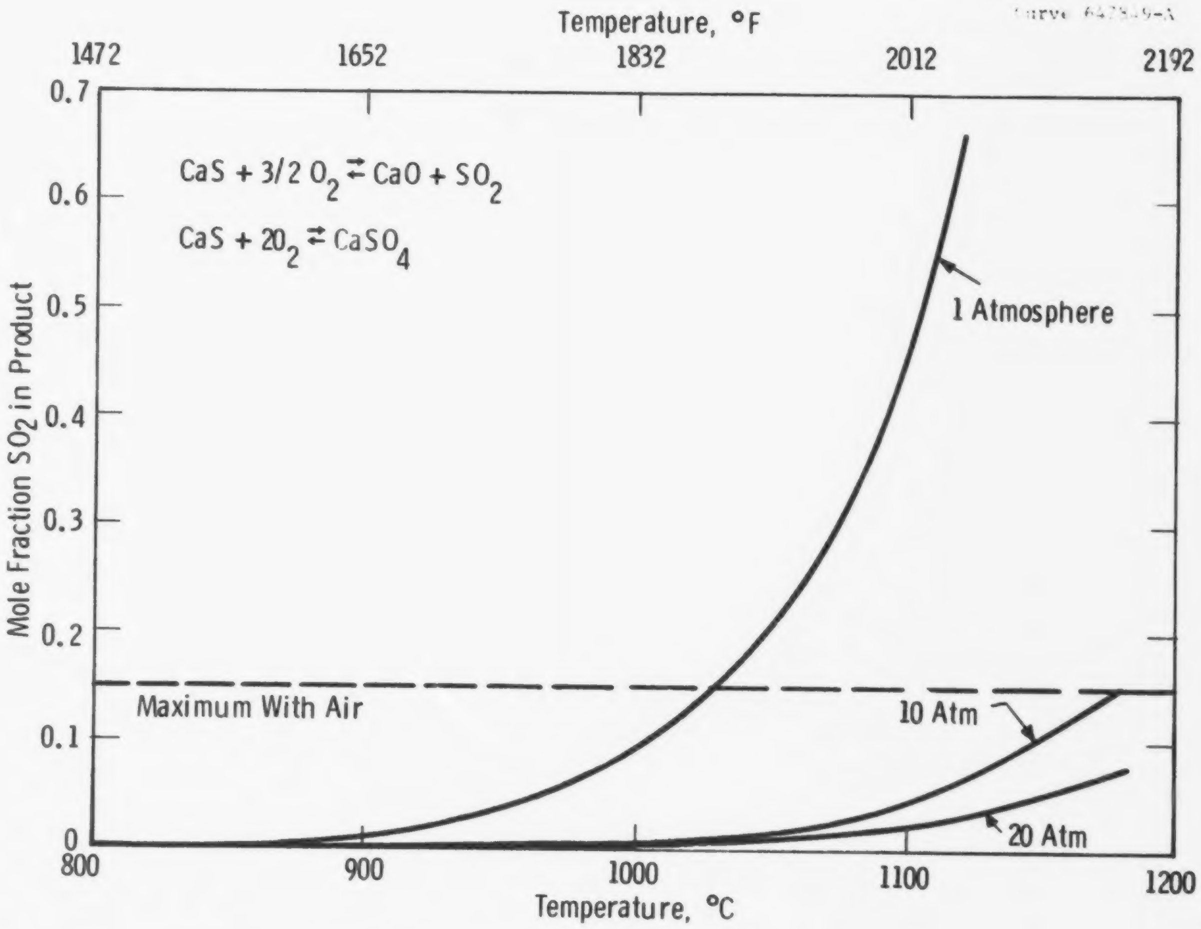


Figure N-1—SO_2 in Regenerator or Sulfate Generator

Curve 647850-B

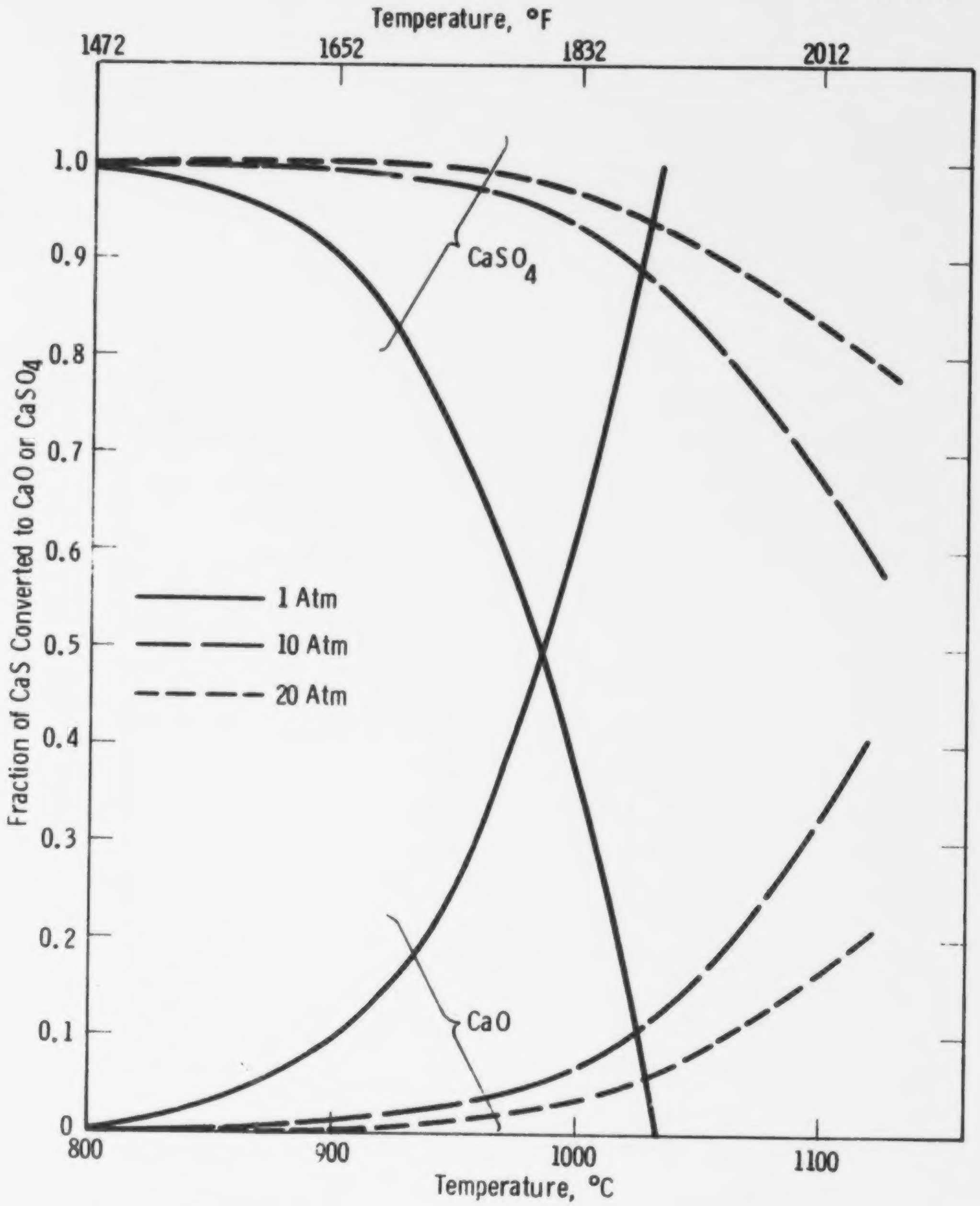


Fig. N-2—Solid Product from Regenerator or Sulfate Generator two figures indicate that the operating temperature of the regenerator or sulfate generator must be changed from that used in the atmospheric-pressure case. In order to maximize the SO_2 concentration from the regenerator if it is operated by reaction 8 at 20 atmospheres pressure, a regenerator temperature on the order of 2200°F or higher would be required. A temperature this high in the regenerator may not be feasible because of a possible degradation of the lime which would cause loss in lime activity. Thus, at pressures as high as 20 atmospheres, some other lime regeneration scheme such as reaction 10 will probably be required. In order to maximize the conversion of CaS to $CaSO_4$ in the sulfate generator at pressures up to 20 atmospheres, a temperature of ∿ 1700°F may be used which will probably improve the sulfate generator behavior over the atmospheric case by improving the kinetics and reducing the temperature control requirements in the sulfate generator.

It is assumed that particle size, gas residence time, bed lime content, lime utilization, and limestone make-up rate affect sulfur removal in the same manner at pressures up to 20 atmospheres as they were found to do at atmospheric pressure. For a given lime particle size, lime utilization, and limestone make-up rate, the gas residence time needed to give a specific sulfur removal is dependent on the total lime content in the gasifier. Based on the results of Esso (England) atmospheric-pressure studies, the residence time necessary for a given sulfur removal is approximately proportional to the square root of the pressure in the gasifier, for a given gas velocity, particle size, lime utilization, and air/fuel ratio. Thus, although the gasifier diameter is greatly reduced by increased pressure, the lime content must be increased by increasing the bed depth as the square root of the pressure. The validity of this assumption must be tested experimentally, but the order of magnitude of behavior is probably correct.

Table N-1 summarizes the preliminary assumptions made with respect to the behavior of this oil gasification/desulfurization process.

Preliminary evaluation of pressurized oil gasification/desulfurization is based on the specifications in Table N-2. Material and

TABLE N-1

ASSUMED BEHAVIOR OF PRESSURIZED OIL GASIFICATION/DESULFURIZATION

	Regenerative Operation	Once-through Operation
Sulfur Removal	same capabilities as at 1 atm need increased gas residence time	same capability as at 1 atm
Metal Retention	same as at 1 atm -- ∿ 100% V, 20% Na	same as at 1 atm
Carbon Deposition[1]	same behavior as at 1 atm though increase pressure may reduce coking	same behavior as at 1 atm
Fuel Gas Composition	CO/CO_2, H_2, CH_4, altered by pressure	CO/CO_2, H_2, CH_4, altered by pressure
Regenerator	operated by reaction 10	---
Sulfate Generator	---	behavior improved over atmospheric case
Temperature Control Requirements	nearly the same temperature control requirements in the gasifier -- decrease in regenerator requirements	same as at 1 atm in gasifier -- reduced in sulfate generator
Gasifier Temperature	same as 1 atm -- ∿ 1600°F	same as at 1 atm
Regenerator -- Sulfate Generator Temperature	∿ 1100°F	∿ 1500°F-1800°F
Limestone Make-up Rate	same as at 1 atm	---
Particle Carry-over[2]	possible reduction depending on choice of operating conditions	possible reduction compared with atmospheric case
Operating Velocity[2]	same range of velocities apply as at 1 atm	same velocity range applies
Bed Depth	$\alpha\ p^{1/2}$ for same sulfur removal	$\alpha\ p^{1/2}$
Gas Residence Time	$\alpha\ p^{1/2}$	$\alpha\ p^{1/2}$
Quality of Fluidization	improved uniformity and mixing	improved uniformity and mixing
Vessel Cross-Section[3]	α 1/P	α 1/P

[1] Carbon deposition, or coking, may be controlled by the air/fuel ratio, location of fuel injection, injection of steam or water, and possibly by air preheat.

[2] Fluidization velocities may be reduced in the pressurized situation since vessel compactness is not as critical as in the atmospheric case. This velocity reduction will lead to reduced particle carry-over. The minimum fluidization velocity and terminal velocity is only slightly affected by pressure at pressures up to 20 atm.

[3] The gasifier cross-sectional area is also affected by the operating velocity, the air/fuel ratio, and the temperature control method.

TABLE N-2
SPECIFICATIONS

Plant Electrical Capacity	250 MW
Capacity Factor	80%
Turndown Ratio	4/1
Number of Gasifier Modules	2
Modes of Operation	regenerative or once-through
Gasifier Pressure	20 atm
Pressure of Gas Turbine	15 atm to turbines
Sulfur Removal	90-95%
Residual Oil	3 wt % sulfur
Gasifier Temperature	1600°F
Regenerator Temperature	1100°F
Sulfate Generator Temperature	1700°F
Air/Fuel Ratios	14 (minimum) - 30% of stoichiometric
Lime Particle Diameters	500-2000 μ average diameter
Regenerative Lime Utilization	10%
Once-through Lime Utilization	35%
Fluidization Velocity	$\geq 2.5\ U_{mf}$, $< U_t$
Limestone Make-up Rate	1 mole CaO/mole sulfur fed
Gasifier Temperature Control	stack gas recycle, steam or water injection
Regenerator Temperature Control	lime circulation rate
Sulfate Generator Temperature Control	excess air circulation (∿ 200% excess)
Plant Heat Rate	9000 Btu/kWh (assumed for purposes of material balances)

energy balances around the gasifier are assumed to be identical with those calculated for atmospheric gasification, though temperature control requirements will be slightly different because of changes in regeneration scheme and heat generation rates in the pressurized case. A plant size of 250 MW with 2 gasifier modules is based on a single gasifier per gas-turbine/steam-turbine combustion. The number of gasifier modules could be increased or decreased, depending on turndown behavior, reliability, etc. A gas-turbine pressure of 15 atmospheres has been chosen, based on projections for gas-turbine conditions. Gas-turbine pressure of 10 atmospheres has also been considered. The gasifier pressure of 20 atmospheres is based on an estimate of 1 atmosphere pressure drop through the gasifier and about 4 atmospheres pressure drop for control purposes. This large pressure drop between the gasifier and gas turbine is extremely conservative. Other specifications are based on the process considerations discussed with regard to the atmospheric-pressure results.

An important factor in determining the best method of gasifier temperature control and the most reasonable air/fuel ratio at which to operate is the system thermal efficiency. The fractional loss in efficiency due to gasifier temperature control requirements has been estimated and is shown in Figure N-3 for stack gas recycle, steam injection, and water injection as a function of the air/fuel ratio. The estimate is based on the gasifier temperature control requirement calculated for atmospheric pressure. The curves for steam injection and water injection are independent of gasifier pressure, but the energy loss for stack gas recycle increases with increased gasifier pressure, since the stack gas must be recompressed to use as a temperature control stream. Operating at the minimum air/fuel ratio of about 14% of stoichiometric is the most efficient air/fuel ratio, but excessive coking may become a problem at this low air/fuel ratio. An air/fuel ratio of 20% should probably not be exceeded, and steam injection should be limited to use at low air/fuel ratios. Water injection appears promising for temperature control in the pressurized operation and may also reduce the coking problem by reaction with deposited carbon.

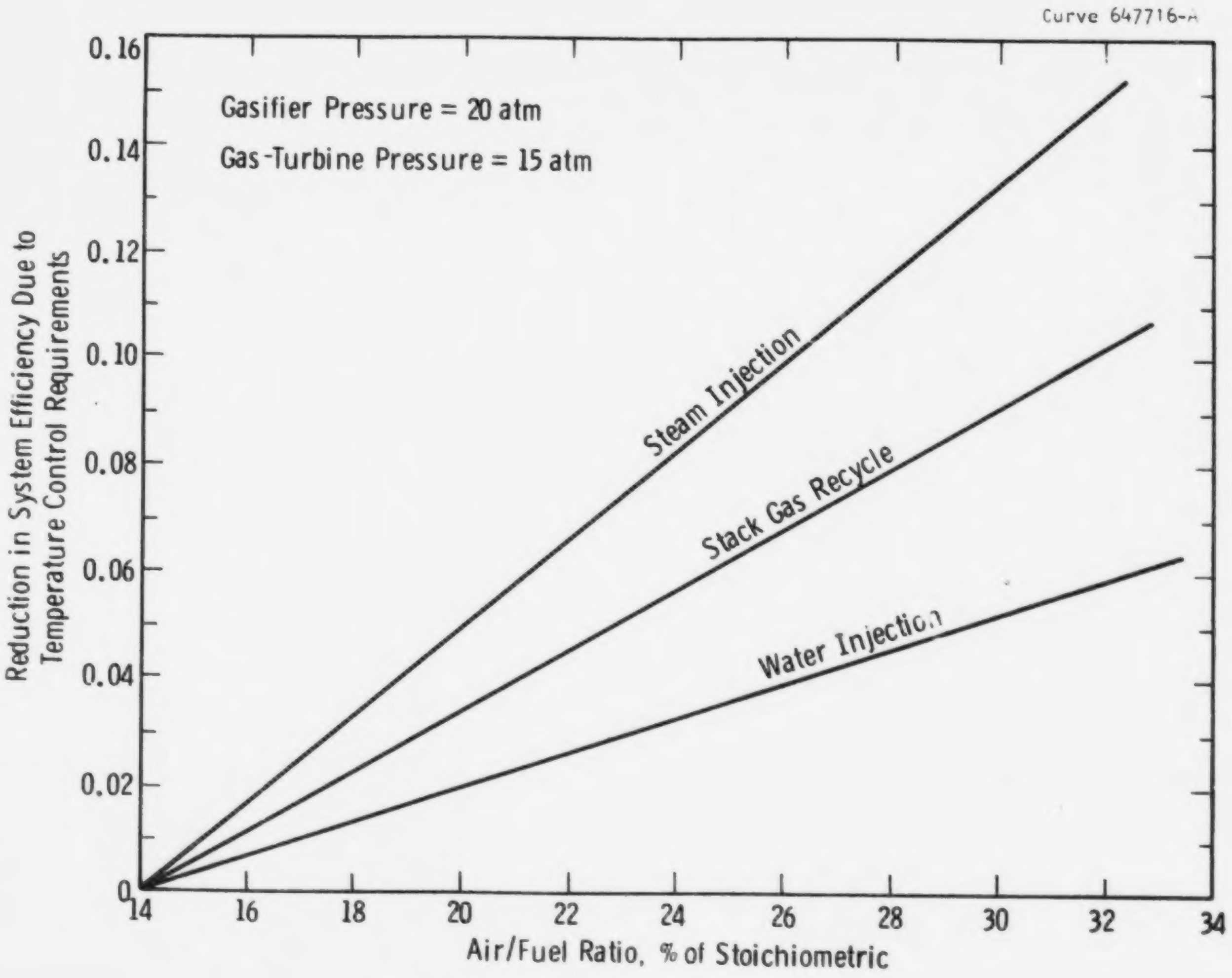


Figure N-3—Energy Loss Due to Temperature Control Requirements

Preliminary designs have been developed for a pressurized oil gasification system. This system will supply fuel gas for a modified PACE 250 MW combined cycle plant. Process alternatives included in these preliminary designs indicate the effect on system cost of variation in air to fuel ratios and of CO_2 source for limestone regeneration.

The block flow diagrams, mass balances, and equipment outline sketches included in this section, based on lean (25% air/fuel) and rich (14% air/fuel) gas production, provide the basis for cost evaluation of the pressurized oil gasification - combined cycle plant combination.

The mass balances tabulated below present flows for one of the two gas turbines in a PACE combined cycle plant. Option 1 utilizes a portion of the fuel gas in a combustor to provide carbon dioxide for limestone regeneration. Option 2 utilizes a portion of the flue gas exiting the combined cycle plant heat recovery boilers to provide carbon dioxide for limestone regeneration. Projected fuel gas compositions are shown in Table N-5.

The block flow diagrams, containing stream numbers and letters referring to mass balance quantities, present major equipment requirements for each process option. These diagrams represent the basic dolomite acceptor gasification scheme, with gravity flow of acceptor to the fluidized regenerator and air transport of regenerated acceptor back to the fluidized reactor.

An additional concept is presented in the last figure in this section. This alternative scheme provides for cooling and particulate removal for the products of combustion of the fuel gas and air before entry to the gas turbine. This route may be required if the hot gas from gasification is shown to carry in vapor form enough ash constituents to give trouble in the gas turbine.

TABLE N-3

125 MW GASIFIER MODULE - REGENERATOR OPTION #1

14% Air/Fuel Ratio (25% Air/Fuel Ratio)

Stream	Temp(°F)	Press (PSIA)	lb./hr	lb./Moles/hr	Composition
1	200	310	6.35×10^4 (6.60×10^4)		85 wt% C, 12 wt% H_2, 3% S
2	200	220	$1.549 \times 10^5 (2.60 \times 10^5)$	$5.34 \times 10^3 (8.96 \times 10^3)$	79% N_2, 21% O_2
3	300	294	$1.287 \times 10^5 (2.36 \times 10^5)$	$4.43 \times 10^3 (8.13 \times 10^3)$	79% N_2, 21% O_2
4	100	310	0 (4.46×10^4)	0 (2.48×10^3)	H_2O
5	300	294	2.62×10^4 (2.37×10^4)	$9.03 \times 10^2 (8.16 \times 10^2)$	79% N_2, 21% O_2
6	1600	280	$1.905 \times 10^5 (3.448 \times 10^5)$	$6.56 \times 10^3 (1.19 \times 10^4)$	CO, CO_2, H_2, H_2O, N_2, CH_4, C_3H_6, etc.
7	1600	275	$1.846 \times 10^5 (3.34 \times 10^5)$	$6.36 \times 10^3 (1.15 \times 10^4)$	"
8	1600	275	$5.94 \times 10^3 (1.078 \times 10^4)$	$2.05 \times 10^2 (3.71 \times 10^2)$	"
9	3500 (3000)	272	3.21×10^4 (3.45×10^4)	$1.10 \times 10^3 (1.18 \times 10^3)$	12.8% CO_2, 10.8% H_2O, 74.6% N_2, 1.8% O (12.0% CO_2, 16.7% H_2O, 69.6% N_2 1.7% O_2
10	250	265	3.21×10^4 (3.45×10^4)	$1.10 \times 10^3 (1.18 \times 10^3)$	"
11	500	25	$1.383 \times 10^4 (1.437 \times 10^4)$	$5.64 \times 10^2 (5.87 \times 10^2)$	25% CO_2, 75% H_2O
12	400	245	8.75×10^3 (9.12×10^3)	$2.82 \times 10^2 (2.93 \times 10^2)$	50% CO_2, 50% H_2O
13	1100	230	1.42×10^4 (1.48×10^4)	$5.95 \times 10^2 (6.18 \times 10^2)$	9% H_2S, H_2O, CO_2, CO, H_2
14	1100	225	1.42×10^4 (1.48×10^4)	$5.95 \times 10^2 (6.18 \times 10^2)$	"
15	100	245	6.86×10^3 (7.13×10^3)	$3.81 \times 10^2 (3.96 \times 10^2)$	H_2O
16 a	800	305	7.77×10^3 (8.10×10^3)	$2.68 \times 10^2 (2.79 \times 10^2)$	79% N_2, 21% O_2
17	300	1250	2.86×10^4 (2.86×10^4)	$1.59 \times 10^2 (1.59 \times 10^2)$	H_2O
18	572	1250	2.86×10^4 (2.86×10^4)	$1.59 \times 10^2 (1.59 \times 10^2)$	H_2O Vapor
19	380	15	5.08×10^3 (5.27×10^3)	$2.82 \times 10^2 (2.93 \times 10^2)$	H_2O
S_1 b	100	294	5.95×10^3 (6.19×10^3)	$5.95 \times 10^1 (6.19 \times 10^1)$	$CaCO_3$
S_2	1600	294	5.21×10^4 (5.42×10^4)	$5.35 \times 10^2 (5.55 \times 10^2)$	10% CaS, 90% $CaCO_3$
S_3	1100	230	4.75×10^4 (4.94×10^4)	$4.75 \times 10^2 (4.94 \times 10^2)$	$CaCO_3$

TABLE N-3 (Continued)

Stream	Temp(°F)	Press (PSIA)	16/hr	16 Moles/hr	Composition
S_4	1100	230	5.79×10^3 (6.02×10^3)	5.79×10^1(6.02×10^1)	$CaCO_3$
S_5 [c]	1600	265	8.85×10^2 (1.57×10^3)	8.85 (1.57×10^1)	$CaCO_3$
S_6	1100	225	1.60×10^2 (1.60×10^2)	1.60 (1.60)	$CaCO_3$
20	250	15		5.36×10^2(5.93×10^2)	N_2, O_2, CO_2, H_2O

---

[a] Solid transport air assumed to carry 5 lb/ft^3.

[b] Carbon deposition of gasifier limestone ~ 0.5 wt % of stone weight.

[c] Solid carryover at rate of 5% of bed per hr.

TABLE N-4

125 MW GASIFIER MODULE - REGENERATOR OPTION #2

14% Air/Fuel Ratio (25% Air/Fuel Ratio)

Stream	Temp(°F)	Press (PSIA)	lb./hr	lb./Moles/hr	Composition
1	200	310	6.35×10^4 (6.60×10^4)		85 wt% C, 12 wt % H_2, 3% S
2	650	220	1.524×10^5 (2.61×10^5)	5.25×10^3 (8.98×10^3)	79% N_2, 21% O_2
3	760	305	1.287×10^5 (2.36×10^5)	4.43×10^3 (8.13×10^3)	"
4	100	310	0 (4.46×10^4)	0 (2.48×10^3)	H_2O
5	1600	280	1.905×10^5 (3.448×10^5)	6.56×10^3 (1.19×10^4)	CO_2, CO, H_2, H_2O, N_2, CH_4, C_2H_6, etc.
6	1600	265	1.905×10^5 (3.448×10^5)	6.56×10^3 (1.19×10^4)	"
7	400	25	1.10×10^5 (1.18×10^5)	3.82×10^2 (4.08×10^3)	3.6% CO_2, 77.8% N_2, 15.5% O_2, 3.1% H_2O
8	280	25	1.383×10^4 (1.437×10^4)	5.64×10^2 (5.87×10^2)	25% CO_2, 75% H_2O
9	220	15	8.75×10^3 (9.12×10^3)	2.82×10^2 (2.95×10^2)	50% CO_2, 50% H_2O
10	500	315	8.75×10^3 (9.12×10^3)	2.82×10^2 (2.93×10^2)	50% CO_2, 50% H_2O
11	1100	305	1.42×10^4 (1.48×10^4)	5.95×10^2 (6.18×10^2)	9% H_2S, H_2O, CO_2, CO, H_2
12	1100	300	1.42×10^4 (1.48×10^4)	5.95×10^2 (6.18×10^2)	"
13	100	320	6.86×10^3 (7.13×10^3)	3.81×10^2 (3.96×10^2)	H_2O
14	760	305	2.37×10^4 (2.47×10^4)	8.17×10^2 (8.52×10^2)	79% N_2, 21% O_2
15	205	15	5.08×10^3 (5.27×10^3)	2.82×10^2 (2.93×10^2)	H_2O (dissolved CO_2)
S_1	100	294	5.95×10^3 (6.19×10^3)	5.95×10^1 (6.19×10^1)	$CaCO_3$
S_2	1600	294	5.21×10^4 (5.42×10^4)	5.35×10^2 (5.55×10^2)	10% CaS, 90% $CaCO_3$
S_3	1100	305	4.75×10^4 (4.94×10^4)	4.75×10^2 (4.94×10^2)	$CaCO_3$
S_4	1100	305	5.79×10^3 (6.02×10^3)	5.79×10^1 (6.02×10^1)	$CaCO_3$
S_5	1600	245	8.85×10^2 (1.57×10^3)	8.85 (1.57×10^1)	$CaCO_3$
S_6	1100	300	1.60×10^2 (1.60×10^2)	1.60 (1.60)	$CaCO_3$
16	250	15		3.26×10^3 (3.49×10^3)	N_2, O_2, CO_2, H_2O

TABLE N-5

GASIFICATION PRODUCT COMPOSITIONS

	Fluidized Bed Process (Vol %)	
	14 A/F[a]	25 A/F[a]
N_2	50.74	47.74
H_2	0.82	2.64
CO	13.40	7.97
CO_2	6.70	7.97
H_2O	0.00	20.56
CH_4	9.43	6.55
C_2H_4 and Higher	18.86	6.55
H_2S	0.05	0.02
Low Heating Value (hot), Btu/scf	∿500	∿280

[a] A/F = Air to fuel ratio, % of stoichiometric.

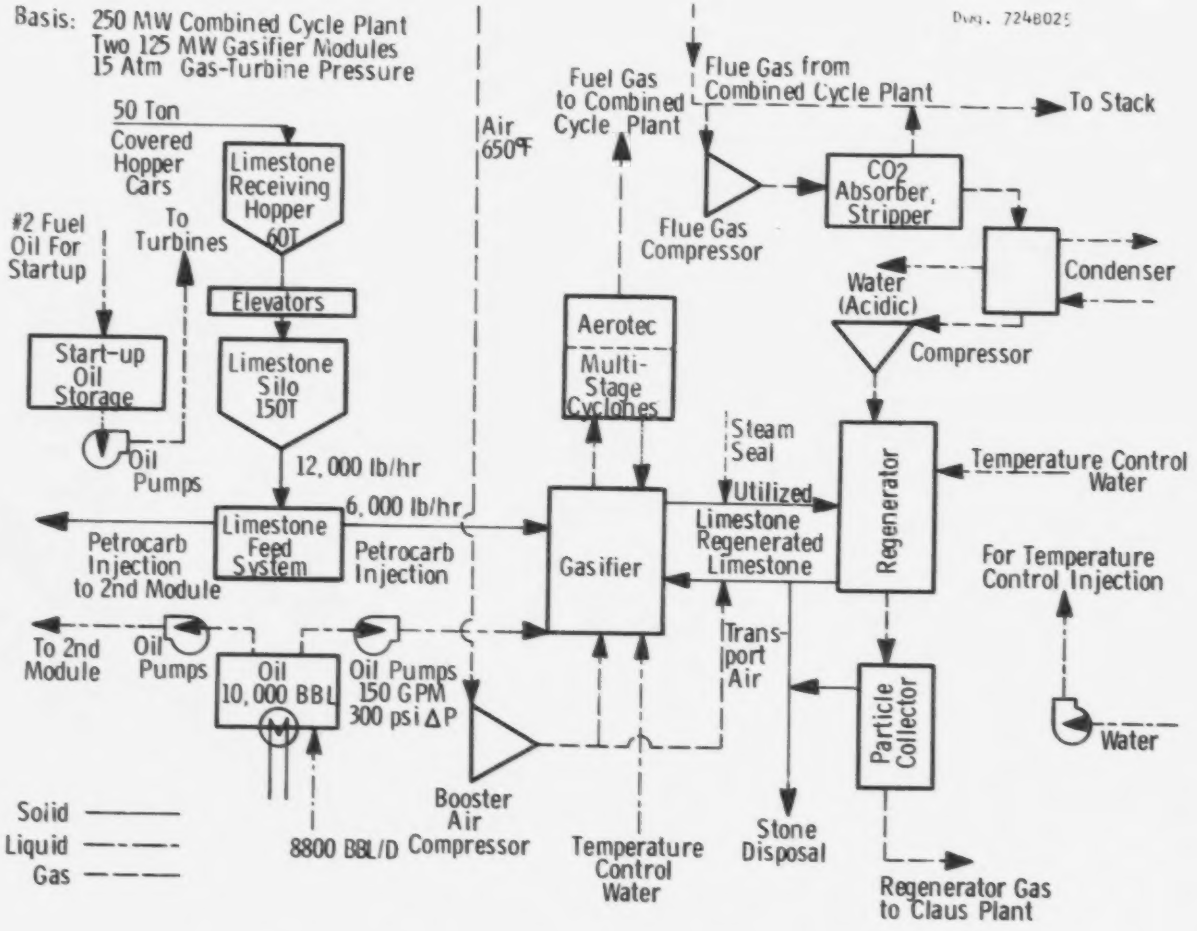


Figure N-4—Pressurized Oil Gasification Plant Flow Diagram

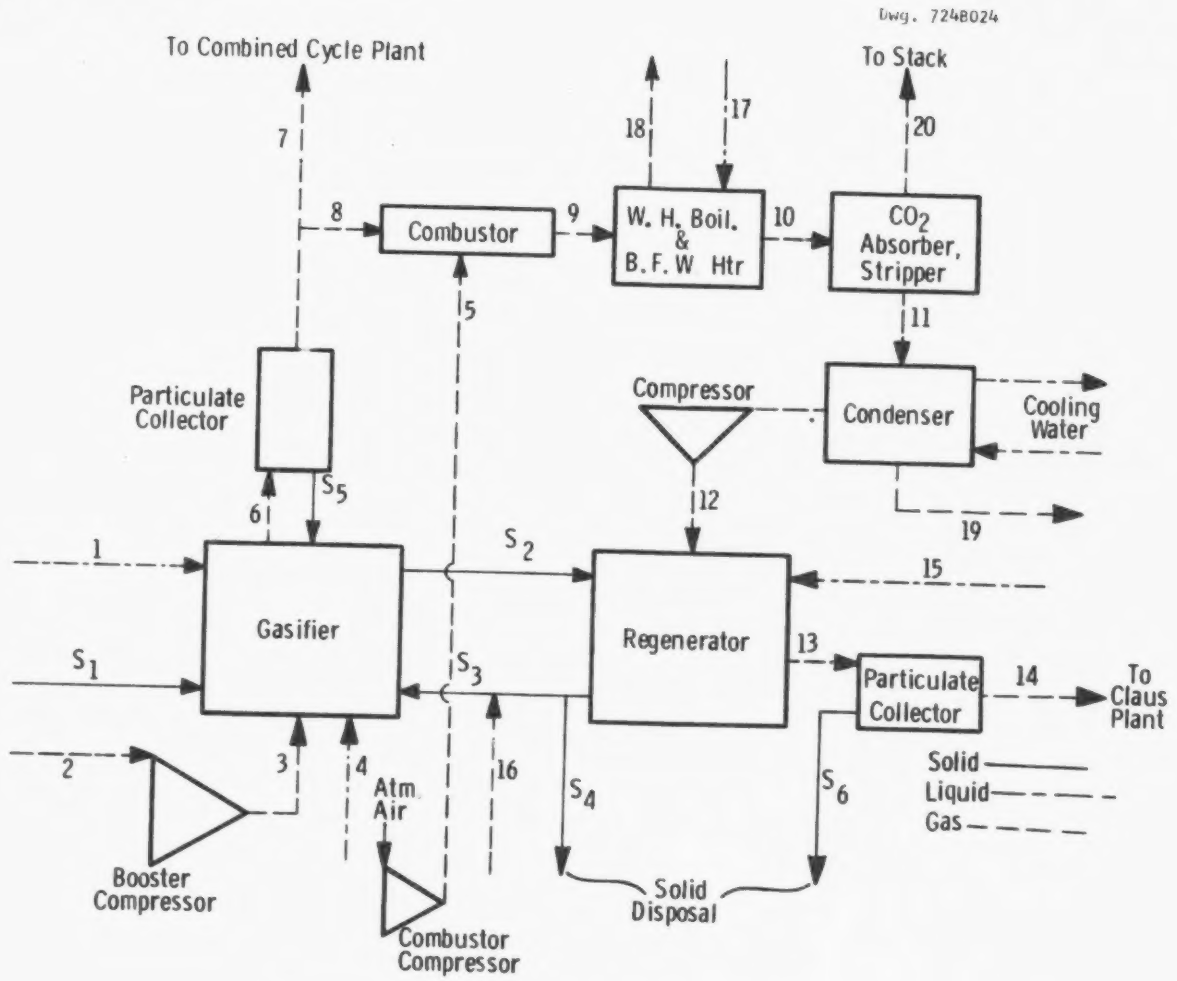
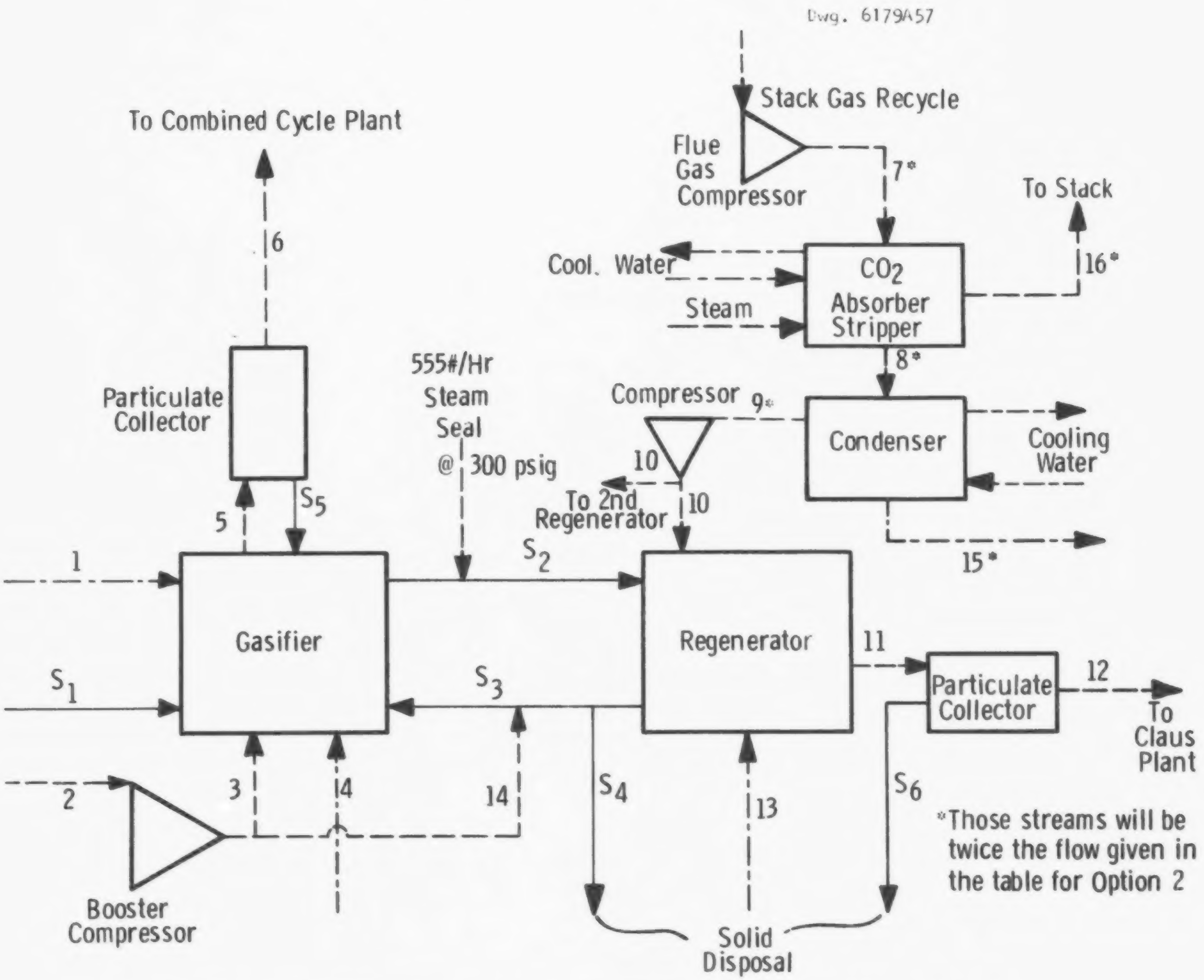


Figure N-5—125 MW Gasifier Module-Regenerator Option 1



*Those streams will be twice the flow given in the table for Option 2

Figure N-6—125 MW Gasifier Module-Regenerator Option 2 (used for estimate)

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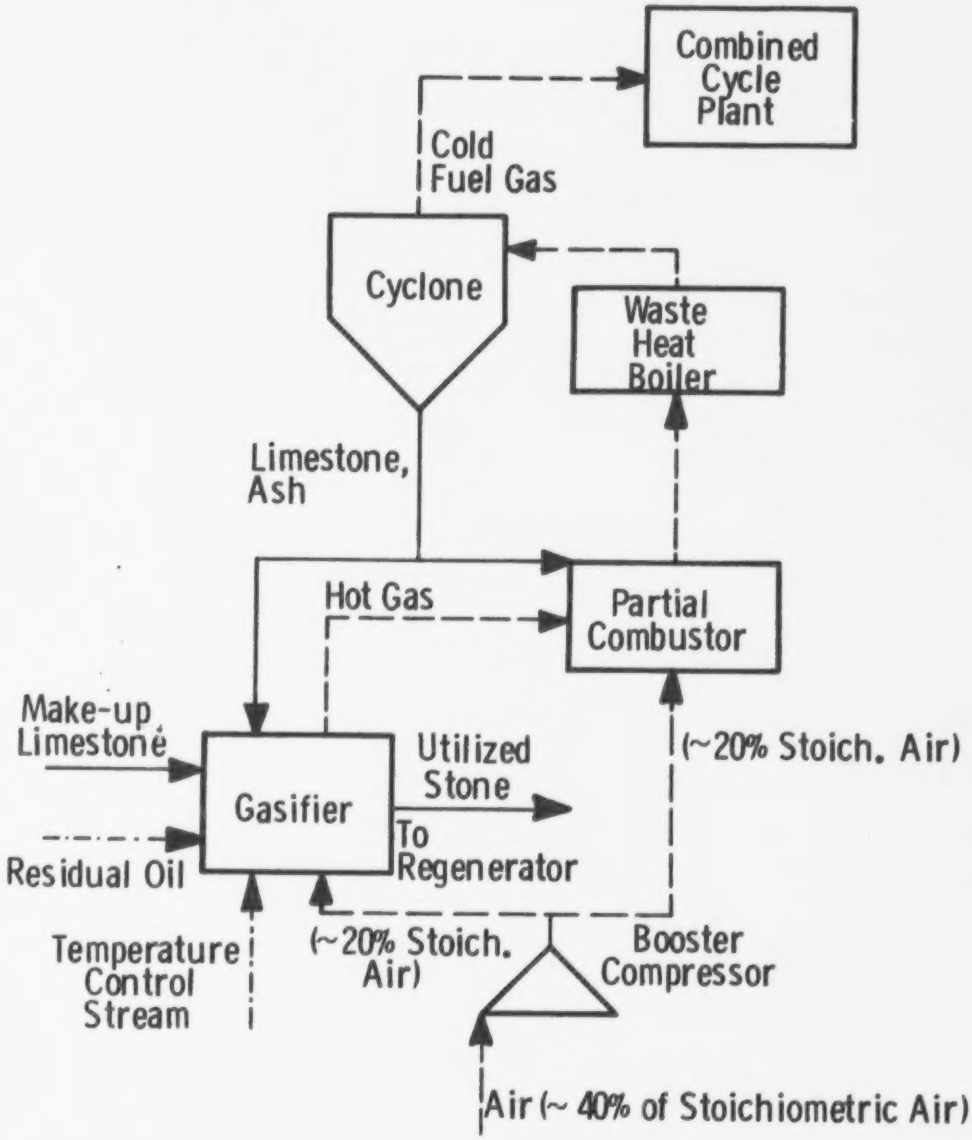

Figure N-7—Alternative Pressurized Oil Gasification Scheme

APPENDIX O
EQUIPMENT DESIGNS

The following considerations are based on a single 125 MW gasifier module and the specifications presented in Table O-1. With the cross-sectional area of the gasifier being inversely proportional to the pressure, Figure O-1 shows the 125 MW gasifier inner diameter as a function of the fluidization velocity and the air/fuel ratio based on previous analysis of atmospheric-pressure gasification. Temperature control is by stack gas recycle or steam injection in the figure, both giving practically identical gasifier diameters. Water injection will yield a slight reduction in diameter.

With the bed depth being approximately proportional to the square root of the pressure for a given sulfur removal, Figures O-2 and O-3 are constructed from atmospheric pressure results to show the necessary static bed depths for 95% sulfur removal in the regenerative and once-through modes of operation. Static bed depth is plotted as a function of fluidization velocity for air/fuel ratios of 14%, 20%, and 30% and for lime particle diameters of 500 μ, 1000 μ, and 2000 μ. For each particle size, the velocity range shown is from 2.5 times the minimum fluidization velocity to the terminal velocity. The factor of 2.5 is used to ensure that the bed will not be defluidized near the distributor plate where the incoming air is cold. In all cases the fuel injection point is assumed to be 20% of the static bed height from the base of the gasifier. This location is important for control of carbon deposition in the gasifier and is based on atmospheric-pressure results. The factors examined in Figures O-1, O-2, and O-3 provide some basis for choosing reasonable operating conditions and design parameters for pressurized operation at 15 atmospheres. Large particles (2000 μ) will allow high velocities up to 10 ft/sec and small gasifier diameters, but large particles at high fluidization velocity will require deep beds (up to 50 ft deep in the once-through operation). The air/fuel ratio also has a dramatic effect on the bed depth with higher air/fuel ratios allowing shallower beds. Once-through operation requires deeper gasifier bed depths because of the higher lime utilization in the once-through case. Although the pressure drop across the bed is of little importance to the system efficiency in the pressurized operation, it is probably not feasible to operate with extremely deep beds due to slugging, segregation of particles, axial temperature gradients in the bed, and increased structural costs. Reasonable limits on vessel inner diameter would be 12 ft (the maximum rail shippable diameter) and a maximum bed depth of about 15 ft, though this limit is not conclusive. These requirements appear to favor fine lime particles ($\sim$ 500μ-1000μ), especially in the once-through operation. Shop fabrication should be possible for most operating conditions, according to Figure O-1.

Gasifier and regenerator vessel designs are shown in Figures O-4, O-5, and O-6. Air/fuel ratios of 14% and 25% of stoichiometric are considered in the figures. Solid circulation between the gasifier and the regenerator is considered in Figure O-7.

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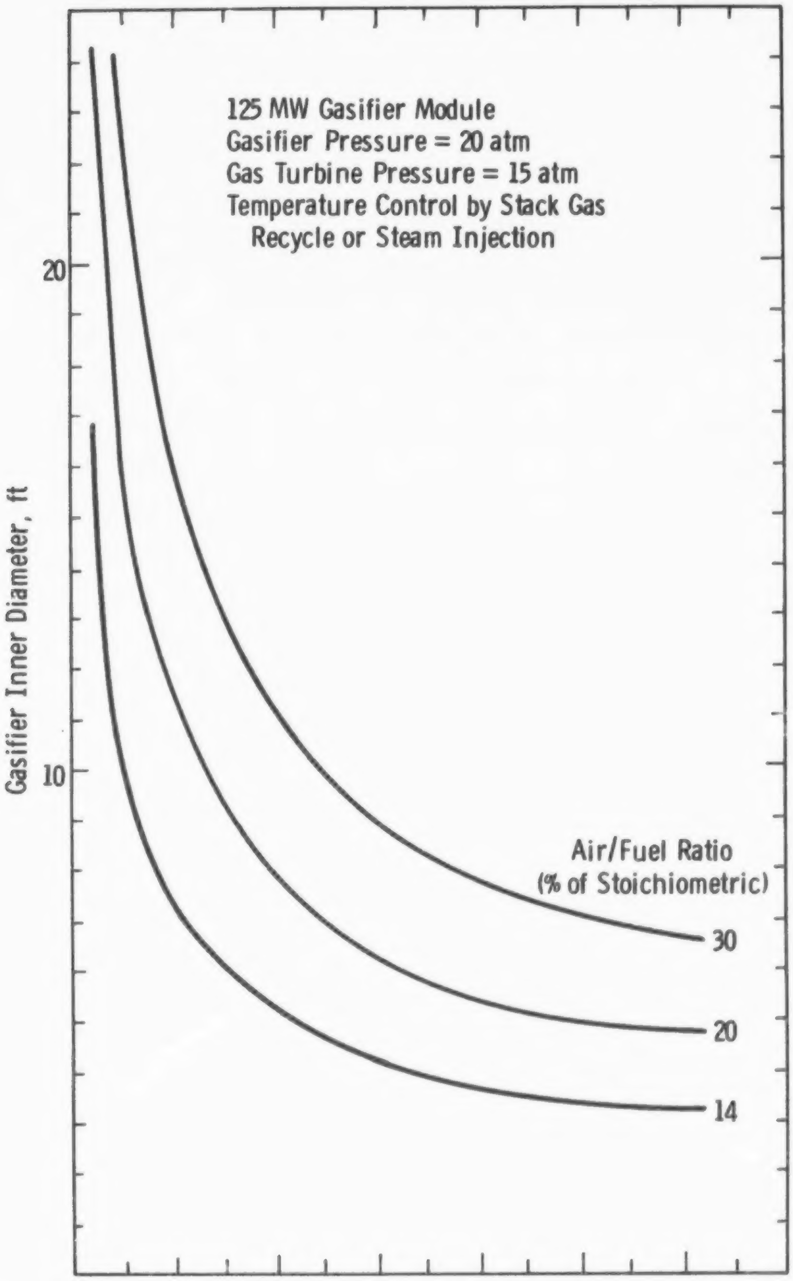

Figure O-1—Gasifier Innder Diameter

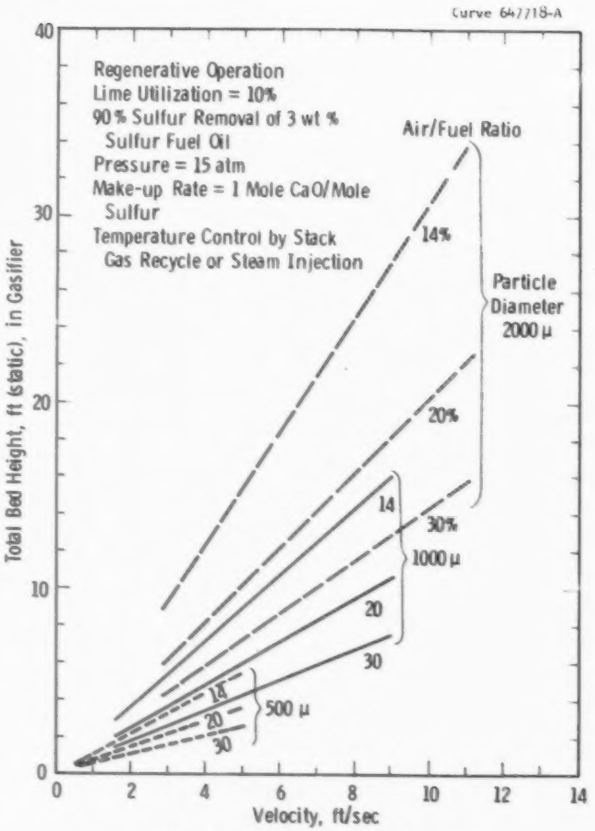

Figure O-2—Regenerative Bed Depth

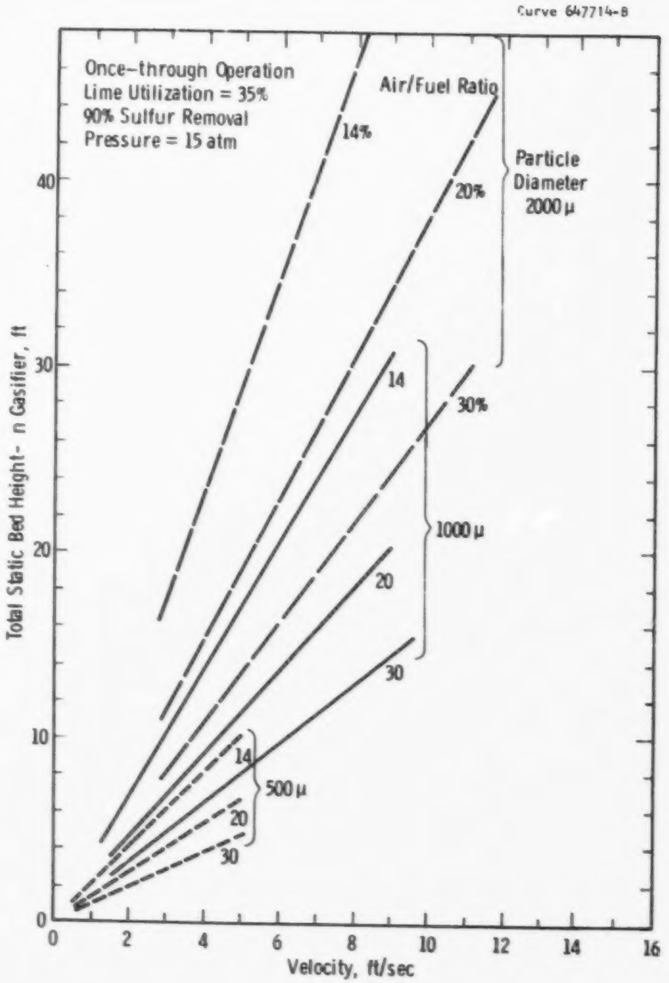

Figure O-3—Once-through Bed Depth

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Basis: 20 Atmosphere Gasifier Pressure; 1000μ Diameter Particles; 5 ft/sec Fluidization Velocity; 125 MW

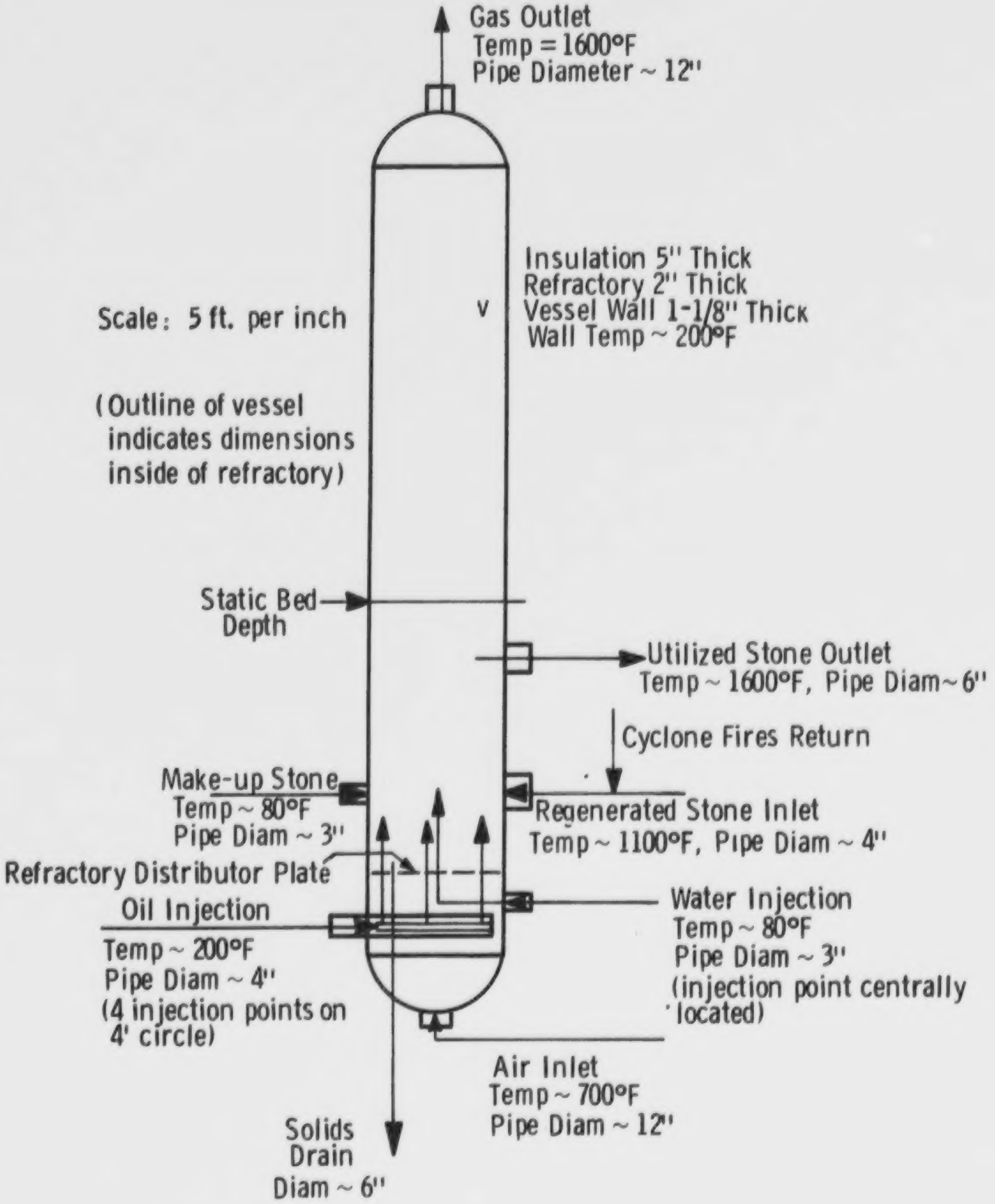

Figure 0-4—Gasifier Vessel at 14% Air/Fuel Ratio

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Basis: 20 Atmosphere Gasifier Pressure; 1000μ Diameter Particles; 5 ft/sec Fluidization Velocity; 125 MW
(Pipe dimensions which are the same as in the 14% air/fuel ratio design are not indicated in this diagram)

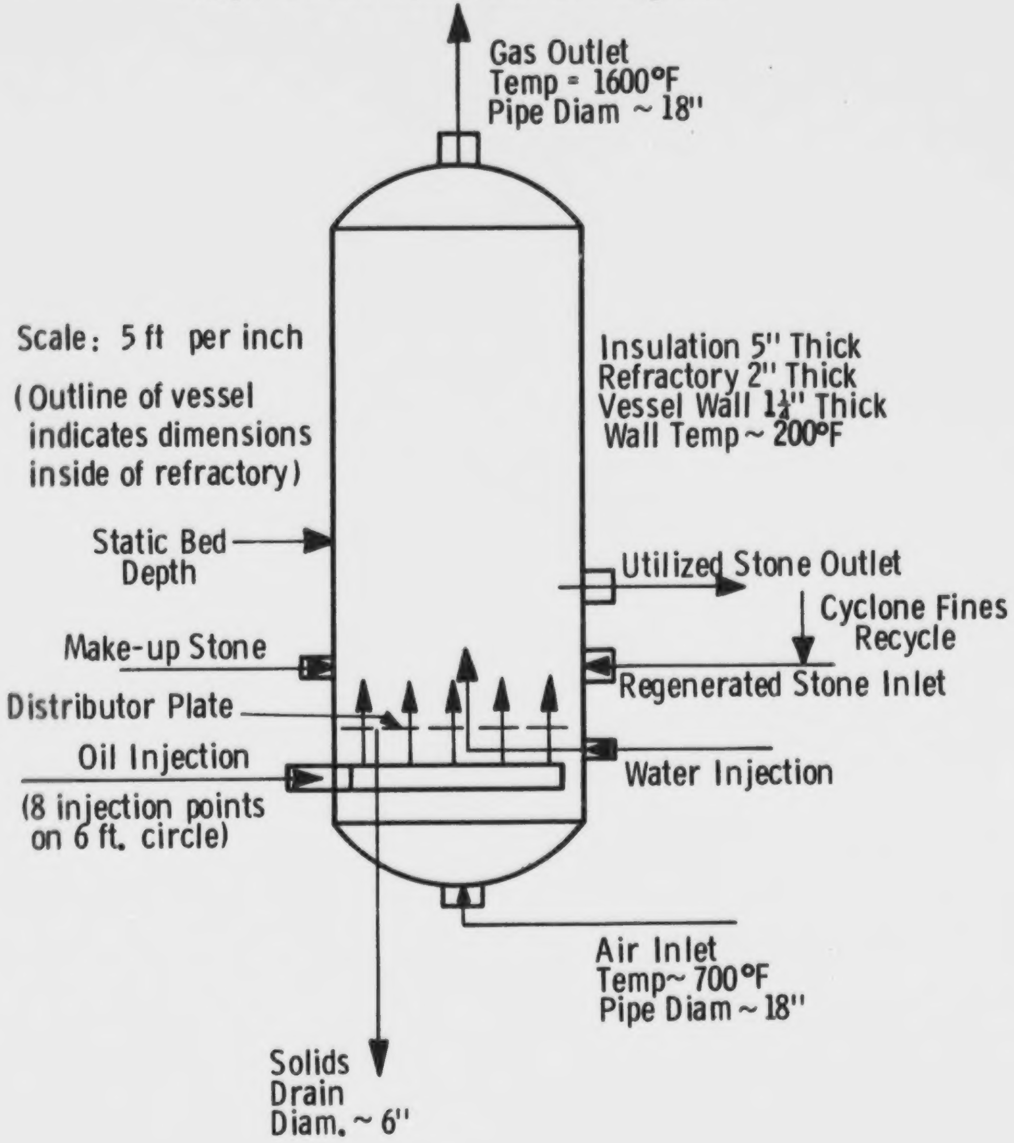

Figure O-5—Gasifier Vessel at 25% Air/Fuel Ratio

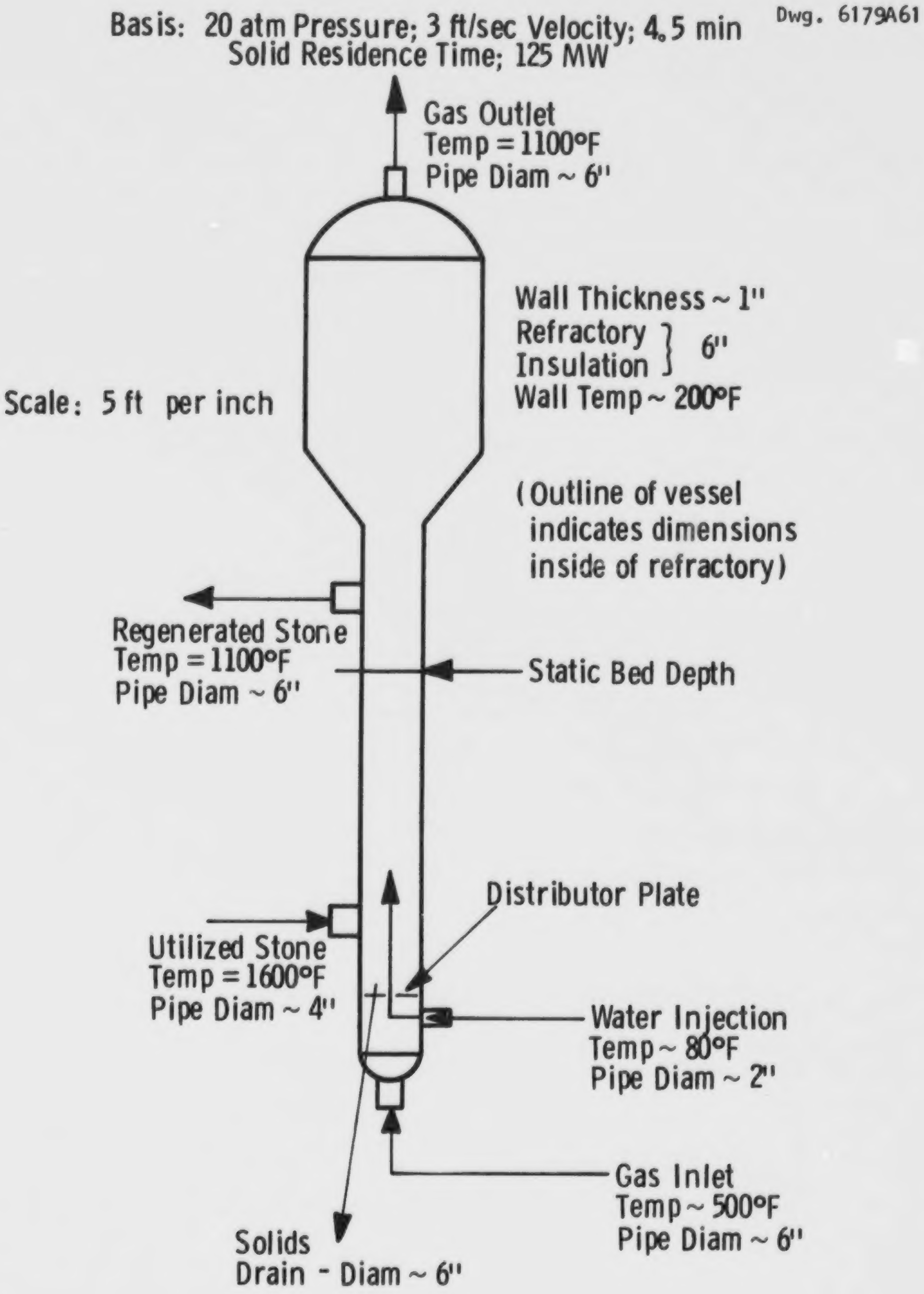

Figure 0-6—Regenerator Vessel Design

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Figures in Parentheses Refer to Gas Pressures in psia

V: Solids Feed Valve

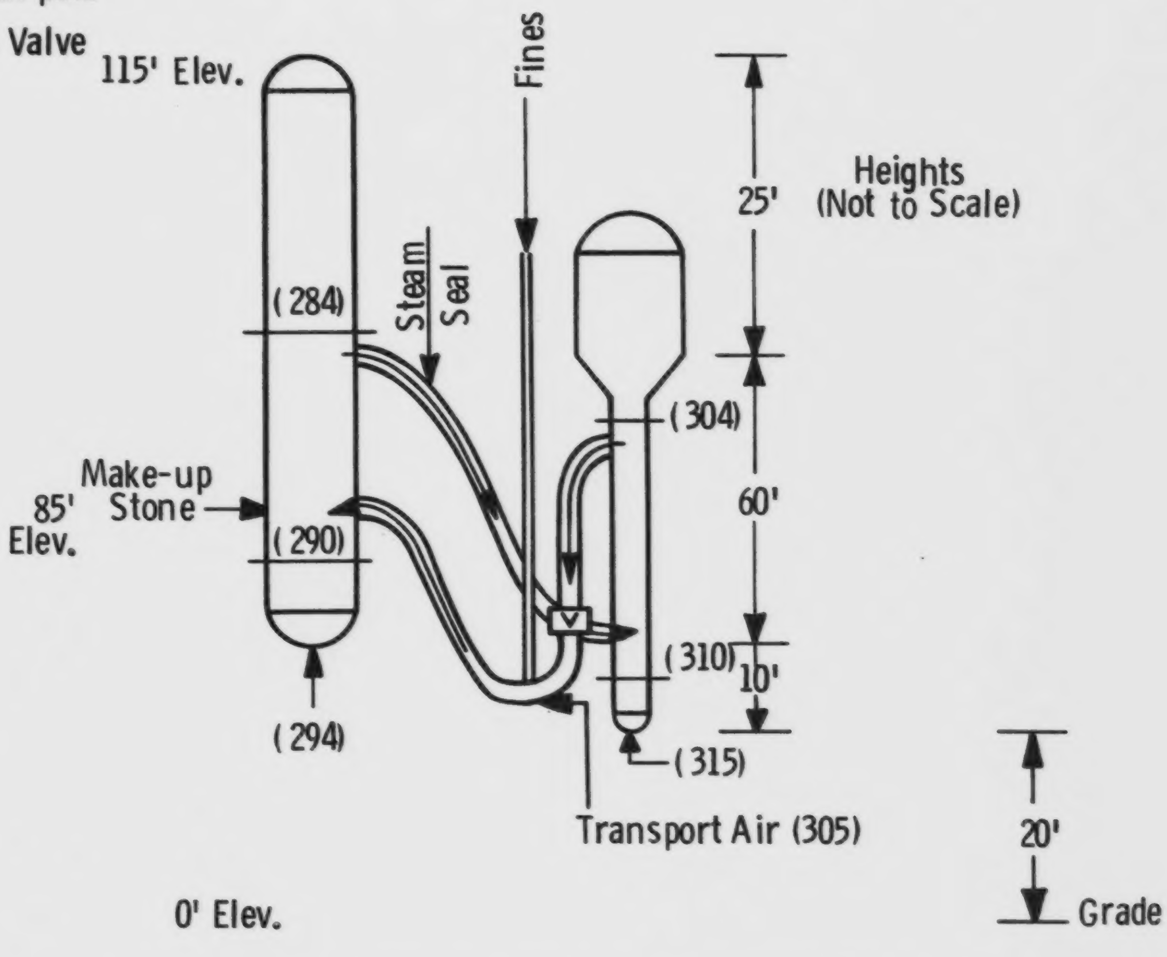


Figure 0-7—Solids Circulation with Regenerator Option 2

APPENDIX P
CAPITAL INVESTMENT

Equipment costs have been estimated from standard cost references appearing in the literature and from vendor quotes. Capital costs shown are total installed costs which reflect both field- and shop-fabrication costs. Each estimate was made based on factoring f.o.b. equipment cost to total installed cost. Each item of major equipment was priced from cost curves derived from published equipment price correlations, modified in the case of pressure vessels and compressors to reflect recent firm price quotations for these two categories of equipment.

The f.o.b. prices of major equipment were totaled; and the total was broken down into percentages for cost of vessels, compressors, mechanical equipment, heat exchange equipment, tanks, and pumps. Appropriate factors were then applied to each category of major equipment to obtain field material costs (pipe, concrete, structural steel, etc.), and field labor costs. Escalation was applied as required to material and labor costs to bring them to mid-1972 levels. The subtotal of all direct costs (equipment, material, and labor) was then factored to provide for field indirect costs, home office engineering, field supervision, freight, and contractor's contingency and fee. The total investment so arrived at is over four times the total f.o.b. equipment cost.

Cost estimates for two 15-atmospheres oil gasification systems were completed; one for 280 Btu/scf turbine fuel gas utilizing a 25% of stoichiometric air to fuel ratio in gasification, and one for 500 Btu/scf turbine fuel gas utilizing a 14% air to fuel ratio. The cost summary for the gasification equipment for a 250 MW combined cycle plant follows:

	500 Btu/scf	280 Btu/scf
A) Equipment Cost		
Exchangers	$ 255,000	$ 255,000
Pressure Vessels	411,000	457,000
Pumps & motors	41,000	58,000
Compressors & motors	620,000	790,000
Tanks	280,000	280,000
Mechanical equipment	546,000	699,000
	$2,153,000	$2,539,000
B) Other Direct Costs		
Construction materials	$1,198,000	$1,376,000
Construction labor	1,424,000	1,662,000
TOTAL A + B Costs	$4,775,000	$5,577,000
C) Indirect Costs	$1,626,000	$1,898,000
D) Claus Plant Subcontract	450,000	450,000
TOTAL COSTS	$6,851,000	$7,925,000
E) Contractors Contingency and Fee	$1,029,000	$1,185,000
TOTAL INVESTMENT	$7,880,000	$9,110,000
Investment based on 250,000 kw gross production from the combined cycle power plant:	$31.50/kw	$36.50/kw

The cost breakdown by sections for the pressurized oil gasification system producing 500 Btu/scf gas is:

	per kw
Air booster compression	$ 3.4
Gasification/sulfur removal	12.9
Limestone regeneration	13.1
Claus plant	2.1
TOTAL	$31.5

This breakdown emphasizes the expense of limestone regeneration. A major portion of the $13.1/kw for regeneration is the production of a CO_2/H_2O reaction gas to produce H_2S from CaS. The reaction gas stream production equipment constituted nearly 90% of the cost for limestone regeneration. The production of a CO_2/H_2O stream is accomplished by absorbing CO_2 from a very dilute (3.6 vol. %) stack gas stream from the combined cycle plant. After stripping, the CO_2/H_2O stream is compressed to limestone regeneration pressure (230 psia).

Costs have been scaled to account for the required product gas pressure as is shown in Figure P-1. Costs are presented as $/MM Btu/hr of fuel oil fed to the gasifier.

The process has been separated into four component systems -- the gasification system, the CO_2/H_2O regeneration system, the booster compressor, and the Claus plant. These four component costs are shown in Figure P-1, in units of $/MM Btu/hr of oil feed (HHV), as a function of the pressure at the gasifier product gas outlet, with the air/fuel ratio as a parameter. Two cases of pressure drops required in the gas turbine combustor and the fuel distribution equipment are considered in the figure -- 8 psi ΔP and 53 psi ΔP. The pressure drop across the gasifier and particle control equipment is assumed to be 9% of the gasifier pressure.

Figure P-1 leads to the following conclusions:

- Reducing the air/fuel ratio from 25% to 14% of stoichiometric reduces the capital investment by about 10%, in units of $/MM Btu/hr.
- For the case of a 10 atmosphere gas turbine, increasing the combustor pressure drop from 8 psi to 53 psi reduces the process cost by about 10% because of the highly pressure sensitive nature of the CO_2/H_2O regeneration scheme.
- A once-through scheme would reduce the cost of the process by more than 40%, neglecting the slight cost increase due to a sulfate generating vessel.
- The cost of the fluidized bed process with an air-blown regenerator would be somewhere between the cost for the CO_2/H_2O regeneration system and the once-through gasification system.

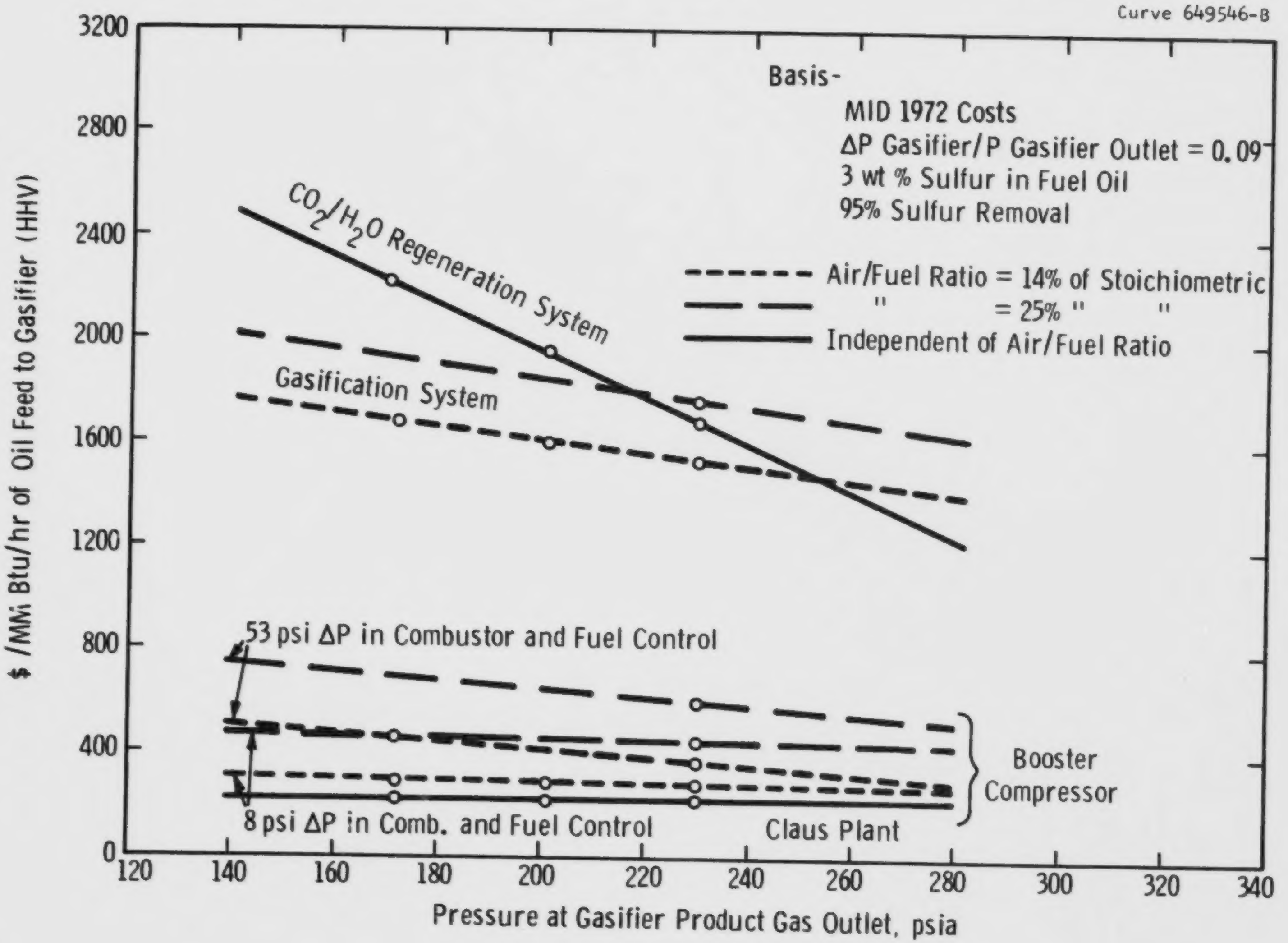


Figure P-1—Fluidized Bed Oil Gasification Capital Investment

APPENDIX Q
PERFORMANCE AND ENERGY COST

INTRODUCTION

The existing Westinghouse proprietory computer programs developed over the last two and one-half years to calculate the performance of gas-turbine and combined cycles were amended to calculate the thermodynamic performance of the PACE-260 combined cycle with either the fluidized bed oil gasification system or the Shell/Texaco oil gasification processes.

This involved both the incorporation of the characteristics of the fuel gas provided by the gasifiers involved and the air or water/steam requirements of the gasifier which could be supplied from PACE plant components (compressor exit, boiler). Gasifier characteristics for two design configurations of fluidized bed gasifiers and the Shell/Texaco process are projected in Appendices A and B. A subroutine for calculating the equipment and operating costs of both the PACE plant and the gasifiers was also constructed and used to calculate capital costs and operating and maintenance costs. Information on gasifier costs for fluidized bed gasifiers and the Shell/Texaco process are summarized in Appendices C and F.

RESULTS

Results for the various gasifier systems include not only a comparison of performance/economics of the systems involved but, in the case of the fluidized bed systems, a study of the effects of total gasifier- fuel gas-supply pressure drop requirements (which determine gasifier design operating pressure) and fuel gas delivery temperature levels (as it affects not only mechanical design of the gas turbine combustion hardware, but costs of gas cleanup systems).

The projected performance of the two fluid bed gasifier systems are compared on Figures Q-1 to Q-5. The assumptions used in these calculations include the following:

- In the case of the Shell/Texaco system the delivered fuel temperature was 100°F. For the fluidized bed gasifiers, the basic delivered fuel temperature was assumed to be 1600°F.
- The design point of the total plant combined cycle-gasifier was at I.S.O. conditions (sea level, 59°F).
- The gasifier pressure drop was assumed to be 9% of the air pressure delivered from the auxiliary booster compressor (which supercharges the extracted air from the main gas-turbine compressor exit). A further pressure drop from the gasifier exit (fuel gas) to the main engine combustor was assumed at 50 psia to account for line loss, fuel control valve pressure drop, and combustor fuel distribution losses. In effect, on the W501B, which delivers 11.3 atmospheres at compressor exit, the total pressure drop was .09 x (11.3 x 14.7 + 50) + 50 or approximately 69.5 psi which is the equivalent of about 4.7 atmospheres. This means the W501B would call for a gasifier operating pressure of 11.3 + 4.7 = 16.0 atmospheres, and the auxiliary compressor would have to deliver a pressure ratio of 16.0/11.3 = 1.4 with a suction temperature of 650°F.

Figures Q-6 to Q-9 show the effect of varying pressure drops on the performance of the PACE plant for two assumed delivered fuel temperatures for the 14% air/fuel and 25% air/fuel ratio fluidized bed gasifiers. The 14% A/F case describes a gasifier which consumes 14% of the stoichiometric fuel and the 25% A/F case describes a gasifier which consumes 25% of the stoichiometric fuel. As such, one can describe the 14% A/F case as an optimum design with minimum process losses, and the 25% A/F case can be described as a nominal design which constitutes a process design with more than adequate margin (or 14% is best estimation, 25% is worst).

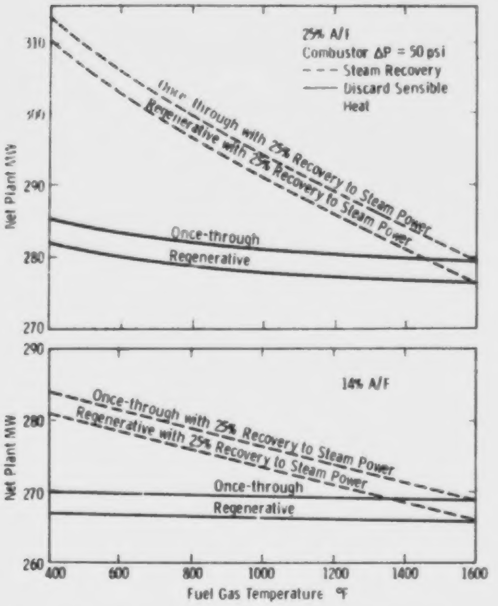


Figure Q-1—Effect of Delivered Fuel Gas Temp. on PACE Plant Power

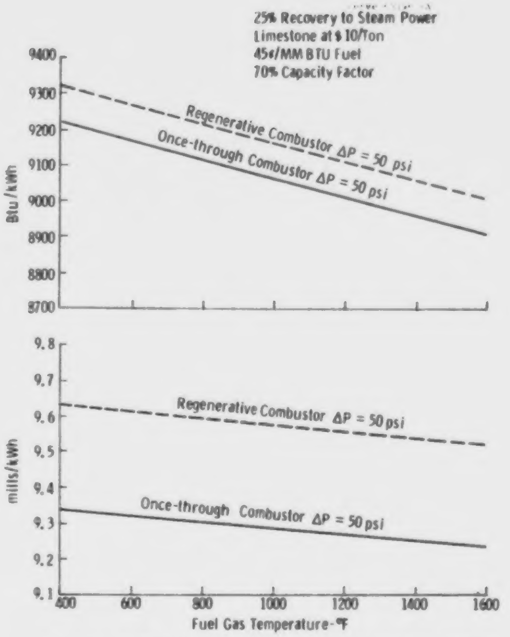


Figure Q-2—Effect of Delivered Fuel Gas Temp. on Heat Rate and Energy Cost - 14% A/F

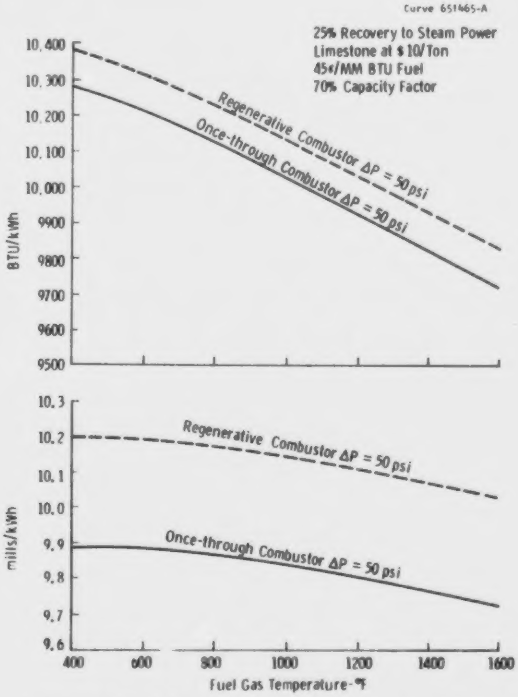


Figure Q-3—Effect of Delivered Fuel Gas Temp. on Heat Rate and Energy Cost

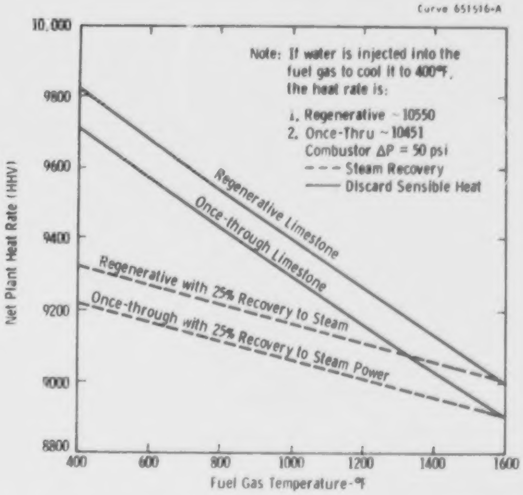


Figure Q-4 –Effect of Delivered Fuel Gas Temp. on PACE Plant Heat Rate 0 14% A/F

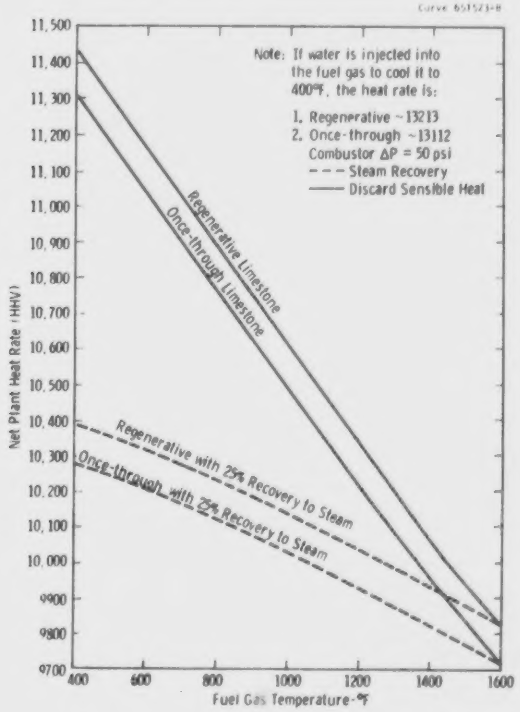


Figure Q-5–Effect of Delivered Fuel Gas Temp. on PACE Plant Heat Rate - 25% A/F

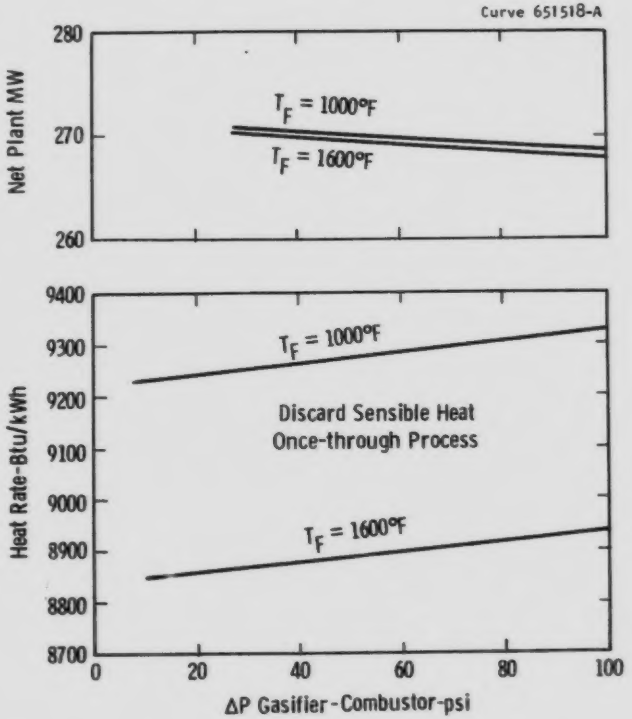


Figure Q-6—Effect of Gasifier-Combustor Pressure Drop-14% A/F

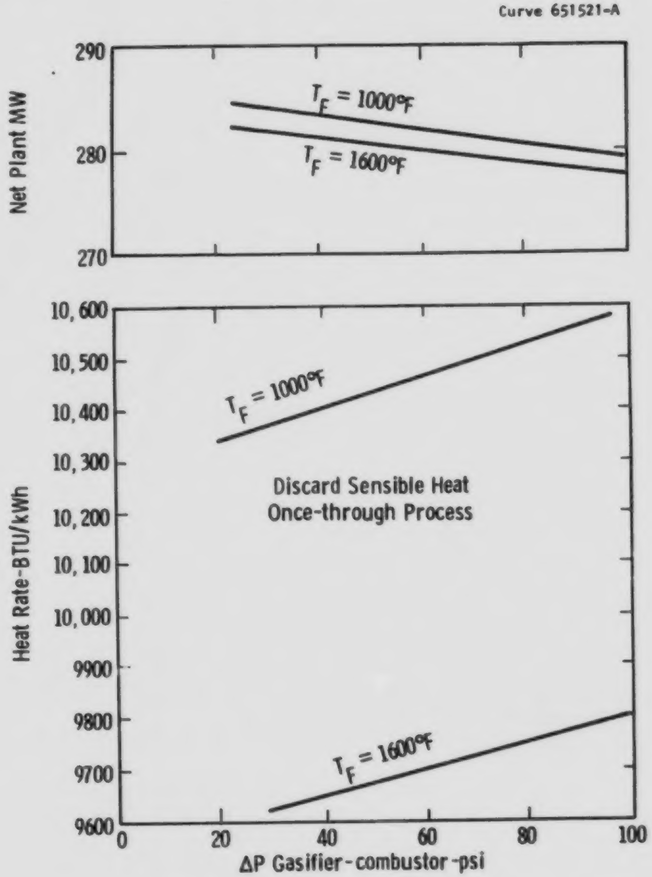
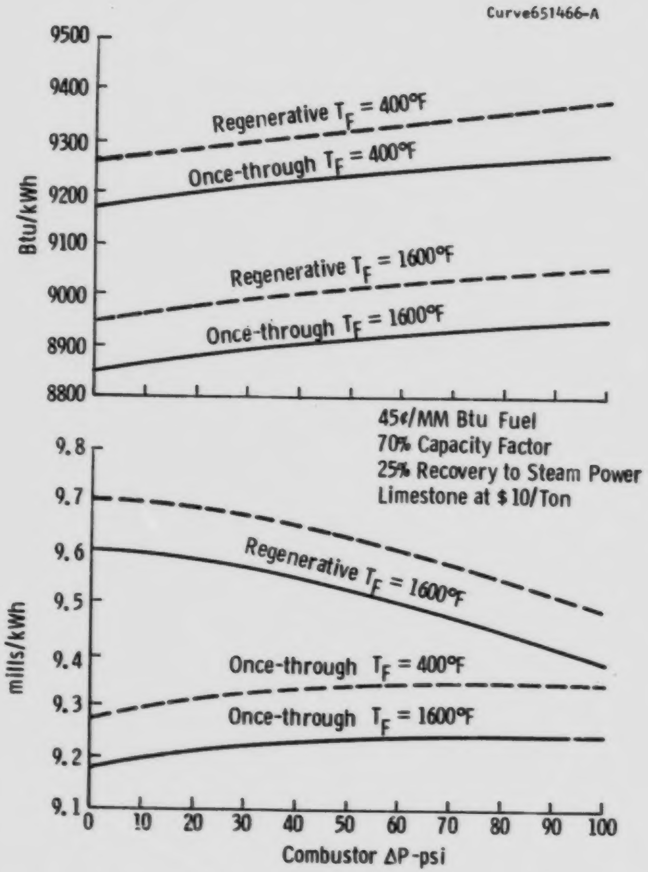
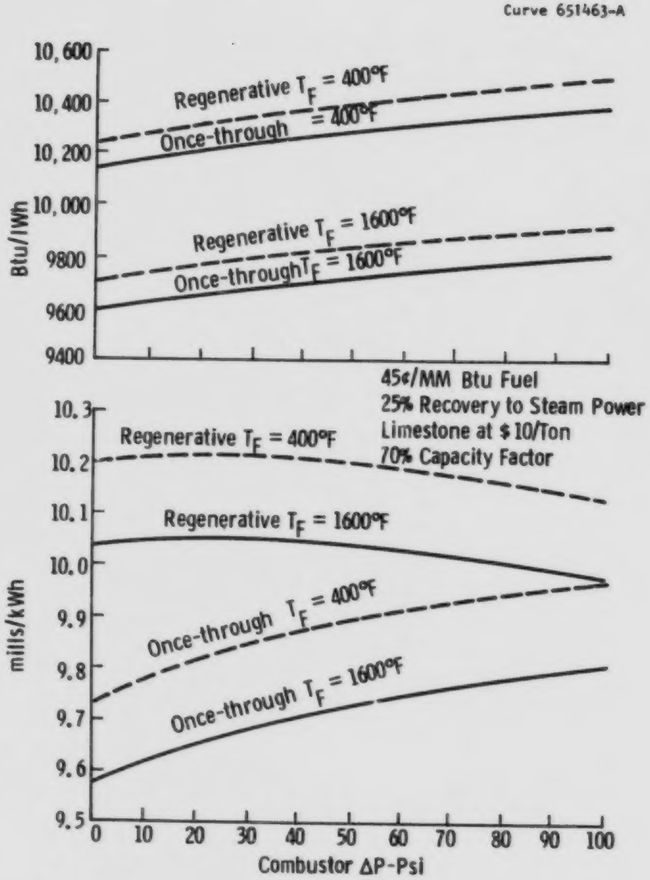


Figure Q-7—Effect of Gasifier-Combustor Pressure Drop-25% A/F



45¢/MM Btu Fuel
70% Capacity Factor
25% Recovery to Steam Power
Limestone at $ 10/Ton

Figure Q-8—Effect of Combustor ΔP on Heat Rate and Energy Cost 14% A/F Case



45¢/MM Btu Fuel
25% Recovery to Steam Power
Limestone at $ 10/Ton
70% Capacity Factor

Figure Q-9—Effect of Combustor ΔP on Heat Rate and Energy Cost 25% A/F Case

Figures Q-10 and Q-11 give the delivered energy costs for the two fluid bed gasifier systems. Optimistic assumptions have been applied to obtain energy costs for the cases of fuel gas temperature reduced below 1600°F. It has been assumed feasible to cool the fuel gas without complications due to high tar content, and the cost assumed for the equipment involved in cooling the gas was low. The feasibility of cooling the fuel should be investigated.

Figures Q-12 and Q-13 represent the performance of the Shell/Texaco processes as a function of the combustor pressure drop.

Figures Q-14 and Q-15 show the effect of the limestone purchase and disposal price on the plant energy cost for the two fluid bed systems. Regenerative and once-through operations are compared.

Figure Q-16 represents the effect of plant capacity factor on the alternative oil-fueled power plant systems. The two fluid bed systems, the Shell/Texaco processes, the PACE plant using distillate fuel, and a conventional plant fired with low-sulfur oil are considered.

Tables Q-1 to Q-4 give performance results for selected processes and operation conditions. Table Q-1 shows plant investments, plant heat rates, and plant power outputs for the selected conditions defined in the table. Table Q-2 shows a breakdown of the energy cost for the processes considered. Tables Q-3 and Q-4 compare the capital investments and energy costs of the selected processes to the capital investment and energy of the PACE plant fired by distillate fuel oil.

CONCLUSIONS

The break-even delta fuel costs indicate that if we take the P65 data on the projected price of high-sulfur residual in the 1975 to 1985 era, which is in the range of 45 to 50¢/MM Btu, the gasification system described (14% A/F case) will compete with other clean liquid fuels (like distillate) if their cost exceeds:

- 60 to 65¢/MM Btu for a once-through gasification system
- 65 to 70¢/MM Btu for a regenerative gasification system.

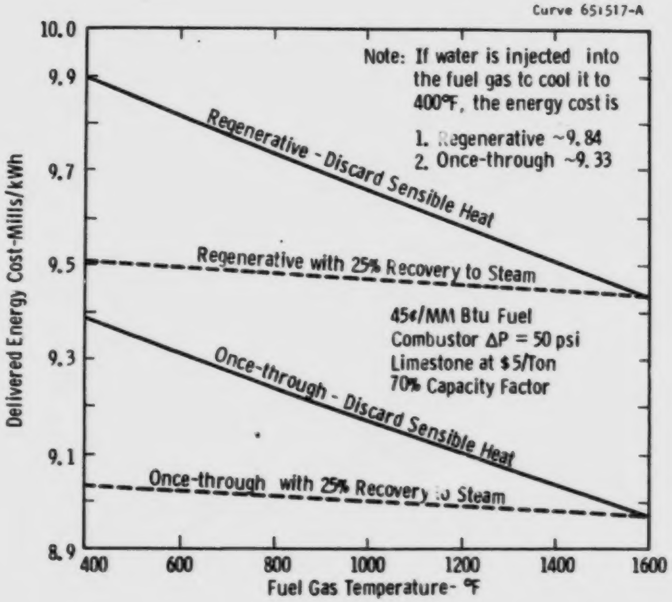


Figure Q-10—Effect of Delivered Fuel Gas Temp. on Energy Cost-14% A/F

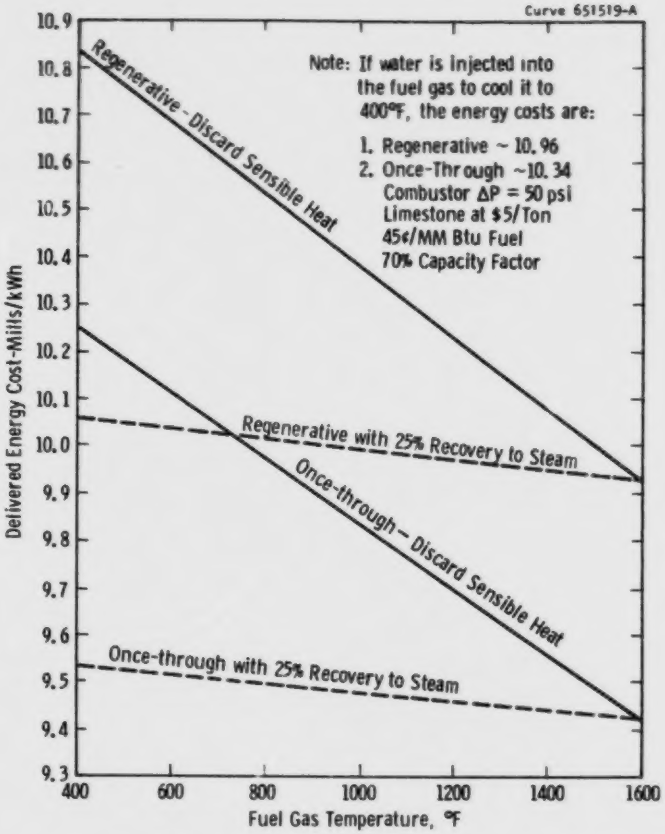


Figure Q-11—Effect of Delivered Fuel Gas Temp. on Energy Cost-25% A/F

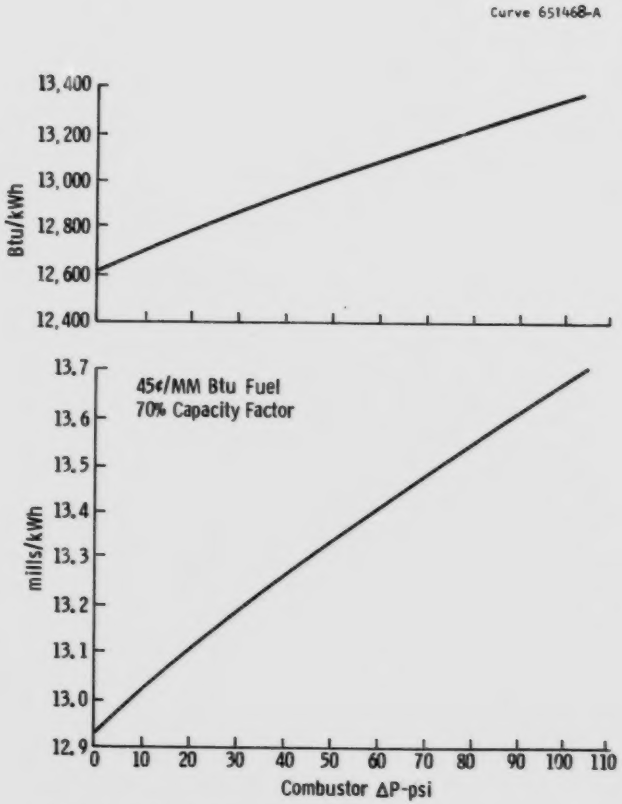


Figure Q-12—Effect of Combustor ΔP on Heat Rate and Energy Cost Shell/ Texaco Oil Gasifier Processes

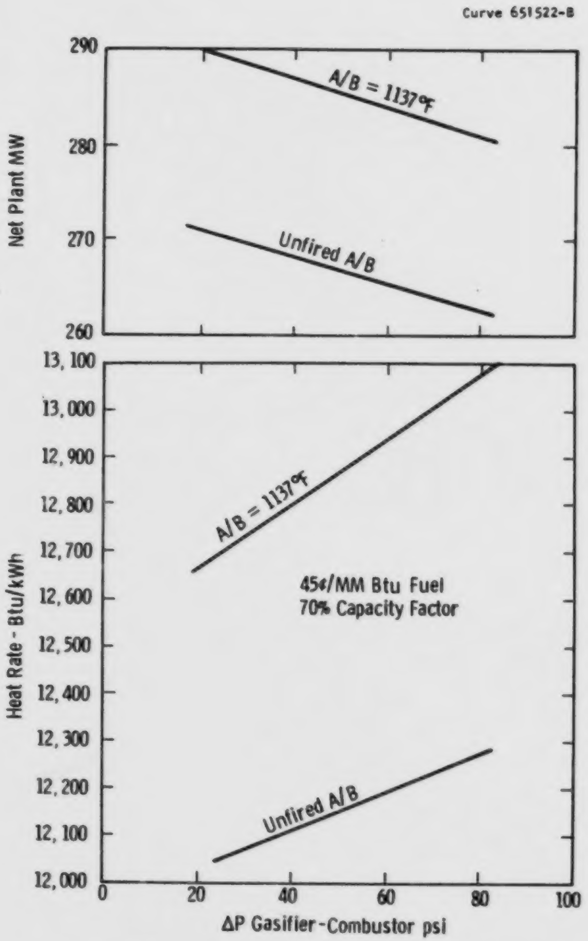


Figure Q-13—Effect of Gasifier-Combustor Pressure Drop on Shell-Texaco Oil Gasifier Processes

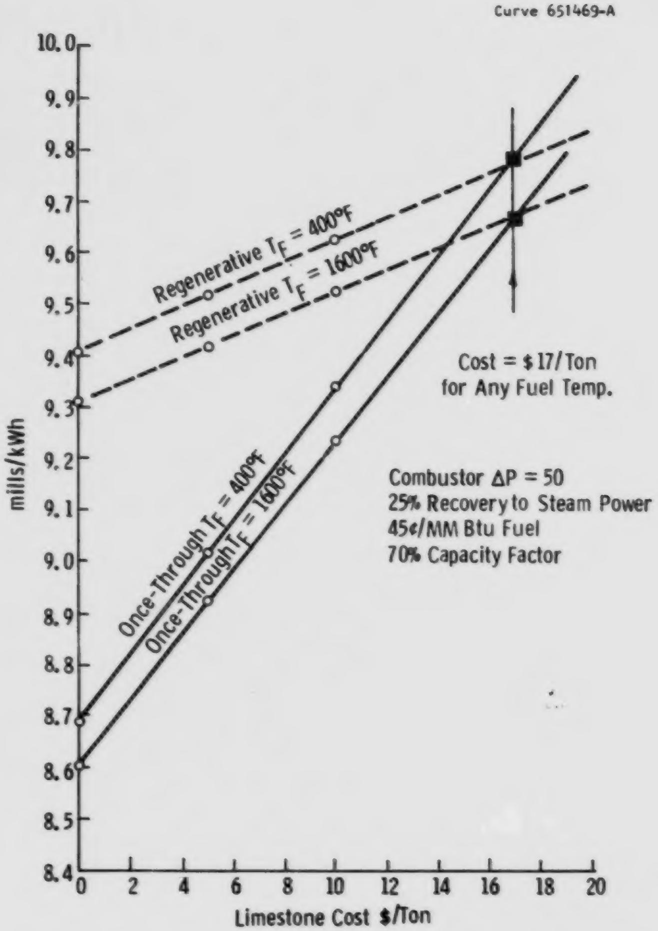


Figure Q-14—Effect of Limestone Cost on Energy Cost-14% A/F Case

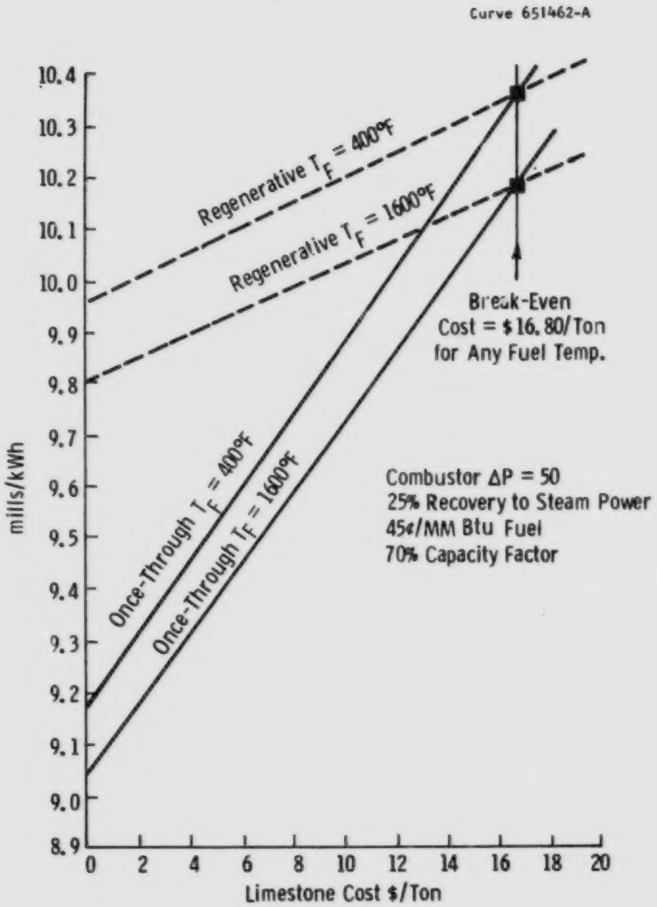


Figure Q-15—Effect of Limestone Cost on Energy Cost-25% A/F Case

Curve 651464-A

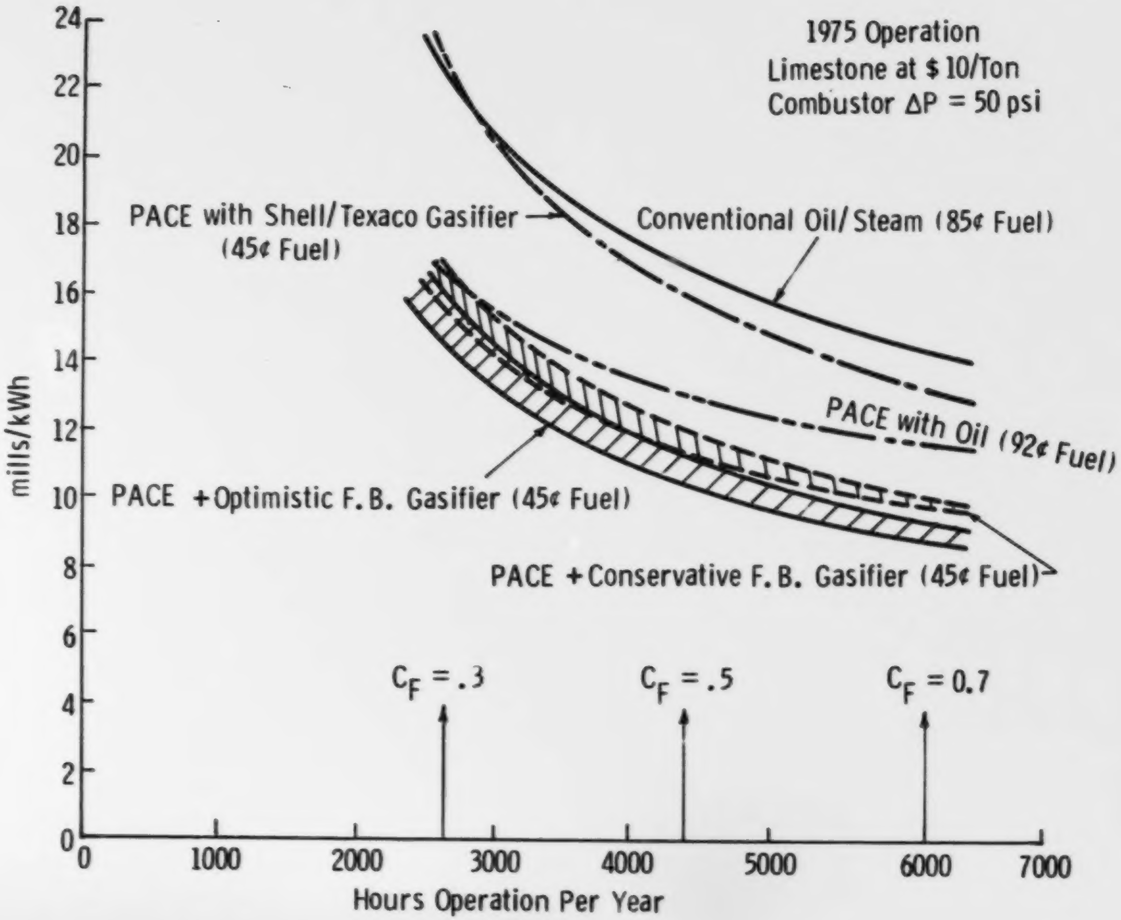


Figure Q-16—Energy Cost Comparison for Various Capacity Factors for ~300 MW Plants

TABLE Q-1

PLANT CHARACTERISTICS

Type	T_F	Type	Lime	CF	\$/kw[a]	Heat Rate[b]	MW	Notes
14% A/F	1600°F	once	\$3/ton	45¢	158.7	8906	268.9	F.B. oil gasifier
14% A/F	1600°F	regen.	\$3/ton	45¢	179.5	9007	265.9	F.B. oil gasifier
25% A/F	1000°F	once	\$10/ton	45¢	162.9	10017	294.2	F.B. oil gasifier
25% A/F	1000°F	regen.	\$10/ton	45¢	185.7	10127	291.0	F.B. oil gasifier
Shell/Texaco	100°F	--	--	45¢	246.7	13005	282.4	oil gasifier
Oil/Steam	--	--	--	85¢	235	9500	300	conventional
Oil/PACE	--	--	--	92¢	137.8	8497	273.6	conventional

[a] Installed cost.

[b] Based on higher heating value.

TABLE Q-2

ENERGY COST DATA FOR 0.7 CAPACITY FACTOR, MILLS/kWh

Type	Capital[a]	Fuel	Plant[b] O&M	Gasifier[c] O&M	Limestone[d] O	Total
14% A/F	3.882	4.008	.520	.196	.189	8.795
14% A/F	4.391	4.053	.520	.347	.064	9.375
25% A/F	3.845	4.508	.520	.245	.710	9.828
25% A/F	4.410	4.557	.520	.414	.239	10.140
Shell/Texaco	6.035	5.852	.520	.928	---	13.335
Oil/Steam	5.749	8.075	.600	---	---	14.424
Oil/PACE (92¢)	3.371	7.820	.520	---	---	11.711
Oil/PACE (45¢)	3.371	3.825	.520	---	---	7.716

[a] 15% capitalization

[b] Based on suggested maintenance contract for GPU proposal for peak continuous operation.

[c] Based on 5% of gasifier total installed cost (including booster compressor system) per year of operation.

[d] Total purchase and disposal cost of limestone.

TABLE Q-3
ADDED COST OF GASIFIER (RELATIVE TO PACE PLANT)

Point	Type	T_F	Type	Plant	PACE	Net Cost
1.	14% A/F	1600°F	Once	158.7	137.8	$ 20.9/KW
2.	14% A/F	1600°F	Regen	179.5	"	$ 41.7/KW
3.	25% A/F	1000°F	Once	157.2	"	$ 19.4/KW
4.	25% A/F	1000°F	Regen	180.3	"	$ 42.5/KW
5.	Shell/Texaco	100°F	---	246.7	"	$108.9/KW

TABLE Q-4
EQUIVALENT COST OF GASIFICATION IN CENTS/MILLION Btu (RELATIVE TO PACE PLANT)

Point	Type	T_F	Type	Delta Fuel Cost[a]
1.	14% A/F	1600°F	Once	12.7
2.	14% A/F	1600°F	Regen	19.5
3.	25% A/F	1000°F	Once	24.9
4.	25% A/F	1000°F	Regen	28.6
5.	Shell/Texaco	100°F	---	66.1

---

[a] Increase in fuel cost over 45¢ which will give equal energy cost for a conventional PACE plant burning distillate. In other words, if a PACE plant used 45+ 12.7 = 57.7¢ fuel, it would have an energy cost equal to the 14% A/F PACE-Gasifier system described as Point 1 in the table.

Overall Conclusions for Cooling the Fluid Bed Fuel Gas

Based on the optimistic assumptions applied for cooling the 1600°F fuel gas:

- For the 14% A/F case the added cost of gas cleanup at 1600°F over cleanup at 400°F cannot exceed:
  - $3.30/kw for the regenerative case
  - $2.43/kw for the once-through case
- For the 25% A/F case the added cost of gas cleanup at 1600°F over cleanup at 400°F cannot exceed:
  - $5.16/kw for the regenerative case
  - $4.64/kw for the once-through case.

TABLE Q-5

BASIS FOR CONCLUSIONS

		14% A/F		25% A/F	
		Regenerative	Once-through	Regenerative	Once-through
1600° Fuel	kw	265,876	268,876	276,158	279,358
	$/kw	179.54	158.75	182.10	159.53
	HTRT	9,007	8,906	9,828	9,716
	E.C.	9.431	8.963	9.932	9.422
400° + Recovery	kw	281,208	284,208	310,380	313,580
	$/kw	176.58	155.25	176.13	152.66
	HTRT	9,323	9,224	10,387	10,281
	E.C.	9.512	9.035	10.059	9.536
Max. Clean-up (1600-400)		$929,000	$833,000	$1,603,000	$1,454,000
$/kw		3.30	2.93	5.16	4.64

Since, by 1975, the projected cost of distillate or other low-sulfur (0.3%) liquid fuels (capable of being used in gas turbines) is projected to be in the 90¢ to 100¢/MM Btu range, a PACE system using the fluidized bed oil gasifier described will have a competitive margin of 30 to 35¢/MM Btu for a once-through system and 25 to 30¢ for a regenerative system.

APPENDIX R

SHELL AND TEXACO PROCESS DATA

- Meeting Reports with Shell and Texaco
- Energy Cost Projections for the Shell Process

SHELL DEVELOPMENT COMPANY

May 3, 1972

Those present:

Shell Development Co. - E. Gordon Foster, Mgr., Process Engr., Licensing
- James A. Sykes, Jr., Process Design, Licensing

Office of Air Programs - S. Rakes

Westinghouse - D. H. Archer
- D. L. Keairns
- W. L. Wright

Summary

The Shell Gasification Process is described in several publications [Syn Gas from Heavy Fuels, Hydrocarbon Processing 12-71, p. 113-117; Partial Oxidation Grows Stronger in U.S., OGJ, 9-6-71].

The discussion was used to clarify and extend our understanding of the process and its capabilities. The following comments were made on the process.

- Air-blown systems have been operated at 100 to 400 psi.
- Maximum capacity of present gasifiers is 40×10^6 scf of $H_2 + CO_2$ per day. The H_2 and CO_2 represent ∿ 40 vol. % of the gas. The heating value is ∿ 125 Btu/scf with air. Vessel diameter is 8 ft O.D.
- Maximum capacity of future gasifiers was considered unlimited (?). They recommend the next scale-up be a 170×10^6 scf/day unit - 125 Btu/scf gas.
- Waste heat boiler accepts ∿ 2500°F gas which contains residual carbon (equivalent to 2% of feed) and no tars.

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The gas velocity in the boiler tubes is selected to prevent solids build-up.

- Gasifier temperature: They operate to maximize production of CO and H_2 and avoid tar formation. They were not sure how low a temperature they could operate and avoid tar formation.
- Gasifier residence time: They use a 5 to 6 sec gas residence time. This is presumably to maximize H_2 and CO. They have not tried going to lower residence times, which would be possible for power generation application since we do not have to maximize H_2 and CO.
- Ash: Some ash collects in bottom of gasifier. They believe ∿ 50% of vanadium, Na and Ni drop out with the ash.
- Maintenance: They shut down every 1 to 1-1/2 years for inspection of refractory and clean-out. They report 95% availability.
- Advantages:
  - All organic nitrogen is removed as N_2; thus if NO_x formation is primarily from organic nitrogen, NO_x can be minimized
  - Demonstrated H_2S removal process.
- Economics: They have capital and operating costs for a 10 atms pressure unit.

Conclusions

- The process is operating and warrants an evaluation for combined cycle power generation.
- Shell will provide preliminary cost information for a gasification system providing fuel for a 50 MW gas turbine. The cost estimate will include the gasifier, waste heat boiler, carbon removal system, and the H_2S removal system.

TEXACO DEVELOPMENT CORPORATION

May 24, 1972

Those present:

Texaco Development Corp. - C. P. Marion, Sr. Process Representative
- F. B. Sellers, Asst. Mgr., Process Div.

Westinghouse - W. L. Wright
- D. L. Keairns
- E. J. Vidt

Discussions were held with Texaco personnel to learn the status of the Texaco partial gasification process. Approximately 60 installations are operating to produce synthesis gas from natural gas and oils.

The following technical features were discussed:

- Air-blown units: No air-blown units have been designed and built. Some units have been successfully operated with air, which were used before the oxygen plant was operational.
- Temperature: Gasifier temperature ranges between 2000°F and 2800°F. They felt it may be possible to operate as low as 1500°F.
- Pressure: Most of the units operate between 300 and 500 psi.
- Thermal efficiency: 70 to 75% without steam credit using air.
- Pressure drop: burners $\sim$ 3 atm
  waste heat boiler 1 to 2 atm
  soot scrubbing $\sim$ 1 atm

- Capacity: The largest commercial unit at the 300 to 500 psi range was reported to be at 400 psi. The vessel diameter was 8 ft I.D. and the gas produced was 40 x 10^6 scf/D of CO and H_2 (O_2 blown). They estimated the capacity of the unit air blown would be $\sim$ 20 x 10^6 scf/D with a heating value of 125-150 Btu/scf. The combustion volume is 285 ft^3. They indicated they could increase the capacity by a factor of 2 or 3 by reducing the gas residence time which is not required if a rich CO + H_2 gas is needed. Nominal R.T. with O_2 is 2-4 sec. Greater capacity increases are possible they believe.
- V, Na, Ni, etc.: "All" are removed in the scrubbing system.
- Turndown: 50% is normal, they can go to 5-10% of load.
- Availability: 95%, annual turn around. This is for operating at essentially full load. They did not believe there would be any problem with cycling the plant for load change; however, they did not recommend cycling the plant for on-off operation.
- Carbon is scrubbed from the gas by water wash, is separated from the water by naphtha, and can be recovered from the naphtha in an oil cut that is suitable for fuel.
- Ash constituents of the resid are recovered in blow-down of a water wash associated with the waste heat boiler. The waste heat boiler design, and how the metallic sulfides and oxides get removed, are proprietary.

They estimated the installed cost of an oil gasification system to be $\sim$ \$100/kw for a 100-150 M W_e system. This includes gasifier(s), waste heat boiler, oil preheat, oil pumps, sulfur removal, sulfur recovery, instruments, piping; but no air compressor, main oil storage, or utilities. The cost is an engineering estimate based on the oxygen fired plants. The time to design and construct a system would be 16 to 18 months in the U.S. (12-14 in Japan, 22 in Europe).

Texaco is conducting an experimental program to determine "optimum" operating conditions for their oil gasification system when applied to low Btu gas production for power generation. Eight to nine months of work have been done and they plan to finish the program within a year.

The process is commercial and warrants an evaluation for combined cycle power generation.

ENERGY COST PROJECTIONS FOR THE SHELL PROCESS

The Shell Process energy cost has been computed on two different bases. The assumption of these two bases are listed in the following table:

TABLE R-1
COST BASIS FOR ENERGY COSTS

Combined Cycle Plant Conditions	Shell Basic Advanced Turbine Technology 2200°F, 14 atm	Ⓦ Basis Current Turbine Technology 1800°F, 11.3 atm
Gasification Process Thermal Efficiency	87%	87%[a]
Oil Cost	\$2/bbl (34¢/MM Btu)	45¢/MM Btu
Gasification Process Cost	\$91/kw	\$91/kw[b]
Sulfur Credit	\$10/ton	none
Capacity Factor	100%	70%
Capital Charges	14%	15%
Plant Heat Rate	8960 Btu/kWh	11,000 Btu/kWh[c]

[a] Previous Ⓦ projection was less than 80%.

[b] Previous Ⓦ projection was \$100/kw.

[c] Previously computed to be 13,000 Btu/kWh based on lower thermal efficiency.

The energy cost breakdown is shown for the two bases for a capital investment of \$254/kw.

TABLE R-2

ENERGY COST FOR SHELL PROCESS -- MILLS/kWh

	Shell Basis	Ⓦ Basis
Capital Charges	4.06	6.21
Fuel	3.05	4.95
Sulfur Credit	-0.06	0
Catalysts and Chemicals	0.06	0.06*
Cooling and Boiler Feedwater	0.40	0.40*
Operating and Maintenance	1.75	1.75*
Total Energy Cost	9.26	13.37

*These numbers are taken from the Shell cost projection and do not correspond to previous Ⓦ cost projections.

On the basis of advanced gas turbine efficiency the Ⓦ basis results in an energy cost for the Shell process of 12.46 mills/kWh. The Shell report which uses the above basis for cost calculations is included for completeness.